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The economics and design of desalination by evaporation through porous, water-repellent membranes

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Department:

THE ECONOMICS AND DESIGN OF DESALINATION BY EVAPORATION THROUGH POROUS, WATER-REPELLENT MEMBRANES

BY

KUEI-HSIUNG YANG -1939-

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THESIS

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Approved by

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Electricity required---------- 2330 Kw.

Capital cost---------- $7,586,000.

Capital cost per gallon per day---------- $0.759.

Product cost---------- $0.596/1000 gallons.

Method II

Capacity---------- 10 MGD.

Performance ratio---------- 10.44 lb./1000 BTU.

Maximum brine temperature----- 500°F.

Heat required---------- 333×10^6 BTU/hr.

Electricity required---------- 2330 Kw.

Capital cost---------- $5,831,000.

Capital cost per gallon per day---------- $0.583.

Product cost---------- $0.472/1000 gallons.

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NOMENCLATURE

A = membrane area required, $ft.^2$

A_H = total heat exchanger area required on production basis, $ft.^2$.

A_{HM} = total heat exchanger area required in the evaporator on production basis, $ft.^2$.

A_M = total membrane area required on production basis, $ft.^2$.

a = activity of water component in salt solution.

C_{Al} = bulk concentration of diffusing component A in fluid 1.

C_{HT} = cost of heat, dollars/day.

C_{MB} = cost of membrane, dollars/day.

C_{MG} = cost of product water, dollars per 1000 gal.

C_{OE1} = cost of the main heat exchanger, $.

C_{OE3}, C_{OEX} = cost of the preheat exchanger, $.

C_{OEV} = cost of the membrane evaporator-condenser, $.

C_{OEG} = cost of engine, $.

C_{OHR} = cost of heater, $.

C_{OMO} = cost of motor, $.

C_{OPM} = cost of pump, $.

C_P = specific heat of salt solution or water.

C_{PO} = cost of power, $/day.

C_t = unit cost of heat exchanger tube, $/ft.

D = diffusivity of component A in fluid 1.

D_A = membrane area required in a small section, $ft.^2$.

D_{AB} = binary gas diffusivity for system A-B.

D_C = product water in a small section, lb./hr.

D_T = temperature driving force across the membrane, ^oF.

D_t = change in stream temperature, subscripts s, c, and s_3 refer to evaporating salt solution, condensate, and salt solution being heated within an increment section.

D_{TEX} = temperature driving force for heat transfer, ^{o}F.

E = elevation of boiling point of 7%.

F = total mass transfer, lb.

H_A = enthalpy of saturated water vapor at temperature of salt water, BTU/lb.

h = latent heat of water, BTU/lb.

h_A = heat transfer coefficient for the film of fluid A, BTU/hr. $ft.^2$, ^{o}F.

h_B = heat transfer coefficient for the film of fluid B, BTU/hr. $ft.^2$, ^{o}F.

h_t = total net positive suction head, ft. of water.

K_m = overall mass transfer coefficient, lb./hr. $ft.^2$ in Hg.

k = thermal conductivity of the membrane, BTU hr. ft. ^{o}F.

k_{cl} = liquid-phase mass transfer coefficient in fluid 1.

k_g = vapor-phase mass transfer coefficient.

k_{g1} = vapor-phase mass transfer coefficient.

k_{g2} = vapor-phase mass transfer coefficient.

k_{p1} = boundary layer coefficient of component A in fluid 1.

k_{p2} = boundary layer coefficient of component A in fluid 2.

L = fictitious film thickness.

N = number of heat exchanger tubes in the membrane evaporator-condenser.

N_A = mass transfer flux, lb./hr. $ft.^2$

N_s = specific speed for centrifugal pump.

P_A = vapor pressure of water vapor at the salt water temperature, in Hg.

P_{A1} = vapor pressure of water on the salt water side, in Hg.

P_{A2} = vapor pressure of water on the product water side, in Hg.

$(P_B)ln$ = log-mean partial pressure difference of stagnant gas B.

P_t = vapor pressure of water at temperature, T.

P_E = total perimeter of heat exchanger tubes, ft.

PIE = principal items of equipment, $.

P_M = perimeter of membrane tube, ft.

Q = rate of total heat transfer, BTU/hr.

Q_c = rate of heat transfer by conduction, BTU/hr.

Q_d = rate of heat transfer by condensation, BTU/hr.

Q_H = rate of heat supplied by the heater, BTU/hr.

q = total heat flux, BTU/hr. $ft.^2$

q_c = conductive heat flux across the membrane, BTU/hr. $ft.^2$

q_d = amount of latent heat transferred, BTU/hr. $ft.^2$

R = gas law constant.

T_c = product water temperature at a given point, ^oF.

T_s = evaporating salt water temperature at a given point, ^oF.

T_{s31} = heated salt water temperature leaving the evaporator and entering the heater, ^oF.

T_{s32} = heated salt water temperature entering the evaporator, ^oF.

T_{op} = total operating costs, $/day.

t = time interval, hr.

U = overall heat transfer coefficient for membrane, BTU/hr.$ft.^2$ ^oF.

U_E = overall heat transfer coefficient for heat exchanger, BTU/hr. $ft.^2 {}^oF$.

W_c = product water flow, lb./hr.

W_s = evaporating salt water flow at a given point, lb./hr.

W_{s3} = heated salt water flow, lb./hr.

X = evaporator length required, ft.

X_t = evaporator length required on a production basis, ft.

Z = length of the diffusion path.

π = Total pressure.

Subscript i denotes at the interface.

Subscripts 1 and 2 denote the inlet and the outlet end of evaporating salt water in the evaporator, if not specified otherwise.

I. INTRODUCTION

Previous investigations (6, 18) have shown that saline water conversion can be achieved by evaporation of water through porous, water-repellent membranes. This process uses the pores of a membrane to serve as the vapor spaces for an almost infinite number of stages of flash-type evaporation and has several thermodynamic, economic and operational advantages over other similar distillation processes and membrane processes. It was desired, based on the present knowledge of the necessary membrane properties, to study the economics and design of this method.

The objectives of this study are:

(1) to study two conceptual designs of a 10 million gallons per day (MGD) desalting plant by vaporization through porous, water-repellent membranes, and

(2) to determine the approximate optimum water cost, the corresponding operating conditions, and how the costs vary with conditions.

II. THEORETICAL ANALYSIS AND LITERATURE REVIEW

A. Desalination Processes:

The separation of water or salt from salt solutions requires energy and the second law of thermodynamics provides a basis for the calculation of the absolute minimum energy required by any desalination process.

In any practical process the energy required is related to the potentials causing transport, and will be much larger than the theoretical minimum (7, 22). The principal irreversibilities or inefficiencies of a practical process correspond to the potentials needed to supply energy to the system and the potentials causing transport of water and salt. Potentials other than chemical, electrical or thermal have not been applied successfully to desalination (7). The chemical potential appears as a result of concentration difference in most instances and as a pressure gradient in reverse osmosis. Electrostatic, electromagnetic, gravitational, and other potentials have been shown little promise but may merit investigation (7).

1. Distillation Processes:

In the distillation of sea water, the salts in solution are (at least in the range of the operating temperatures and pressures) essentially nonvolatile. Thus, in principle, pure water only

evaporates to the vapor phase and, by condensing provides pure liquid water as an end product.

The main distillation processes are described below:

(a) Multiple-effect long-tube vertical:

In any distillation operation, the major process cost is in the steam consumed. Therefore, methods of reducing steam consumption (or of increasing economy) are very attractive. Multiple-effect distillation has been used for this purpose.

The sea water is boiled in the first vessel, using steam as the heating medium. The vapor boiled off is not condensed by circulating water, but it is condensed by being used to evaporate vapor from brine in the second vessel. This process is continued in series down the chain of evaporators. In this multiple-effect distillation, each effect is at progressively lower temperatures and pressures. The vapor produced from the final effect is condensed by circulation of cooling water.

(b) Multi-stage flash:

The multi-stage flash system is related to the multiple-effect system but in this case feed is preheated by condensing vapors plus additional heat and then is "flashed" to lower pressures in stages with some evaporation from sensible heat in each stage. This method eliminates internal evaporator heating surface entirely, so that all vapor is obtained by flashing, and heat is recovered by condensing the flashed vapors with feed solutions.

This process has an advantage in simplicity since pumps are required only to supply the raw feed and remove the final concentrate and condensate. However, it requires more heating surface than an evaporator of the same economy and has a high power consumption because of the large volume of feed and concentrate that must be pumped in relation to the condensate output (14).

(c) Vapor recompression:

Another method of increasing steam economy is by recompressing the vapors from the evaporator and feeding them back into the steam chest of the same effect. Vapor is compressed primarily to increase the pressure and temperature so that the heat released on condensation can be returned to the evaporator for further evaporation.

This system encounters several mechanical and economical disadvantages (12):

(i) The compressor is expensive and is subject to higher maintenance costs than is the remainder of the evaporator system.

(ii) In reasonable operating ranges, the temperature driving force obtained is small, about $10^{\circ}F$. As a result the evaporator must be large for any reasonable production rate.

2. Membrane Processes:

Membrane processes (with the exception of the process under study) fall into two major groups, electrodialysis and reverse osmosis.

(a) Electrodialysis:

Electrodialysis is a membrane process which, by using an electric current as a driving force to move salt ions through solution, takes advantage of the property of certain plastic membranes to pass only cations and other membranes to pass only anions, and thus separate the salts.

The amount of electric current required, and resulting demineralization cost, is in proportion to the amount of salt to be removed. Consequently, the process is favored for brackish waters containing considerably less dissolved salts than sea water. If the cost of membranes and equipment can be sufficiently reduced, an electrodialysis process may become economically feasible for demineralizing sea water (22).

(b) Reverse Osmosis:

Reverse osmosis separates the water from the saline solution by a pressure applied to the salt water side of the membrane. If the applied pressure is in excess of its osmotic pressure, fresh water flows through in the opposite direction to normal osmotic flow. Reverse osmosis differs from electrodialysis in that water is removed from salt water than salt from water and that the driving force is pressure rather than electrical potential. This process has long been recognized as one that potentially could be operated at a high thermodynamic efficiency. Short membrane life, low membrane flux and rejection, high energy requirements, high equipment costs, and concentration polarization are the principal

problems that exist and that must be overcome in order for the process to achieve the optimistic expectations (11).

(c) Vapor-gap reverse osmosis:

The vapor-gap reverse osmosis process differs from conventional reverse osmosis process in that the separating medium is not a semipermeable membrane, but a so-called vapor-gap. The gas-filled gap, maintained between two porous barriers or permeable membranes, supports the pressure differential and separates the liquid and the gas phases. This process is based on (water) vapor pressure over a saline solution being raised above the vapor pressure over pure water, by sufficient pressurization of the saline solution. For the process to function as described, a barrier is required between the pure water and the pressured sea water that is permeable to water vapor, but impermeable to the pressurized sea water and the unpressurized fresh water. In Narmco's reports (4, 5), the barrier requirements for both forward and reverse osmosis experiments were met by using hydrophobic fibers.

3. Evaporation Through Porous, Water-Repellent Membrane:

When a sea water and fresh water system separated by porous, water-repellent membrane is at a steady state without mass transport, a temperature potential, corresponding to the elevation in boiling point of the solution, is required. When sea water is at a temperature higher than the above static temperature, a net transfer of pure water occurs by evaporation from the sea water

and condensation into pure water. Thus, desalination is achieved by this evaporation and condensation mechanism. The pressure gradient which results from the temperature difference across the pores of the membrane serves as a driving force to cause transport of vapor from sea water to vapor phase in the pores, through the pores, and to the fresh water liquid by condensation. The phase boundary permits rapid transport of water vapor, but not salt. Transport from liquid to vapor is brought about by a small thermal potential and the separation process is highly efficient.

A pore of the membrane serves as a vapor space in contact with hot saline water and cooler fresh water. A flash evaporator stage may contain only a liquid section, a vapor space, and a condensate section with the necessary inlets and outlets. Similarly, a single vapor-filled pore may act as a flash evaporation stage. In this discussion flash evaporation is considered to be evaporation where heat of vaporization is supplied by sensible heat of the solution. On this basis, Findley (6) has pointed out that a porous membrane could act as an infinite-stage flash evaporation system.

In the distillation processes, the separation from liquid to vapor is considered to be highly efficient. However, large losses of energy are introduced by the temperature differences required to supply the needed energy to the boiling solution by heat transfer, and this is accentuated by formation of scale. The latter (7) is primarily responsible for limiting operating temperatures to about 121°C (250°F).

In most evaporators each effect is at a different pressure, and liquids must be maintained at suitable pressures and levels by pumps, valves, and controllers. However, in the infinite-stage flash evaporation through porous media system suggested, the liquids involved could be at any convenient pressure higher than the highest boiling pressure, and the liquid section would be completely filled. No changes in pressure are required for liquids, the pressure and temperature of evaporation are limited only by the liquid temperatures and the presence of non-condensables in the pores. Thus, the present process has the following advantages over other distillation processes:

(i) Less pumping and control equipment is required.

(ii) Scale formation is unlikely against a non-wettable surface of the hydrophobic membrane.

(iii) Operation at higher temperatures may be possible.

(iv) Corrosion in the evaporator becomes avoidable by special arrangement of equipment.

(v) The porous, hydrophobic membrane serving as an infinite stage flash system permits this process to operate at a high efficiency.

The major disadvantages of this process are that it introduces increased resistance to mass transfer and allows some heat to be conducted to the coolant without evaporation.

Electrodialysis and reverse osmosis utilizes membranes that are selectively permeable to either ions or water. Gilliam and McCoy (7) stated that the required energy for the processes is

supplied efficiently, so that the losses encountered are largely in the separation processes themselves. Both involve fairly large potential losses because of the gradients needed to effect reasonable rates of mass transfer to and from the membrane surfaces and through the membranes. For example, in reverse osmosis, the separation process is inefficient, inasmuch as a large additional pressure (above 22 atm. required for equilibrium with sea water) is essential to obtain acceptable transport rates (7). In electrodialysis the resulting demineralization cost is in proportion to the amount of salt to be removed, and application is now restricted to brackish waters (22). Again, Gilliam and McCoy (7) have pointed out that either process may cause concentration gradients near the membrane which tend to oppose the transfer.

But the process suggested here has the equivalent high efficiency in separation from liquid to vapor as the distillation processes, and it does not meet such large energy losses in separation as the other membrane processes do.

Although there are several potential advantages over other processes, the evaluation of the suggested process depends on the following factors:

(i) performance of membrane,

(ii) life, strength and cost of membrane,

(iii) time effect on the performance of membrane,

(iv) auxiliary equipment cost and performance.

B. Mass Transfer Through Porous, Water-Repellent Membranes:

When salt water and pure water are separated by porous, water repellent membranes, the direction of mass transfer is determined by the vapor pressures exerted by both sides. The vapor pressure on the salt water side is a function of temperature and concentration of salt solute, and on the pure water side it is a function of temperature only. At a fixed solute concentration, temperatures on both sides determine the directions of mass transfer the following way:

1. No net mass transfer when salt water is at the temperature where its vapor pressure is the same as the vapor pressure of the pure water at its temperature. Under these conditions the salt water temperature minus the fresh water temperature is the boiling elevation of the salt water.

2. Water on the pure water side evaporates and condenses on the salt water side as in forward osmosis when salt water is less than the boiling point elevation above the fresh water temperature.

3. Water on the salt water side evaporates and condenses on the pure water side when salt water is above the fresh water temperature by more than the boiling point elevation. A net product of water can be attained in this case.

Case (3) serves as the basis for the desalination process under consideration. The mechanism of mass transfer of Case (3) is discussed below:

(a) Diffusion of water from salt solution to the surface of the membrane results from a concentration gradient in the boundary

layer, with a concentration drop $(C_{Al} - C_{Ali})$ across it.

Whitman (2) first suggested that the two film theory could be applied to the mass transfer of this vapor-liquid model and the mass transfer flux can be expressed as

$$N_A = k_{cl} (C_{Al} - C_{Ali}) \tag{2.1}$$

where k_{cl} is the liquid-phase mass transfer coefficient in fluid 1, C_{Al} is the bulk concentration of diffusing component A in fluid 1, C_{Ali} is the interfacial concentration of diffusing component A in fluid 1.

Sherwood (3) stated that separation processes can be based on the development of concentration gradients within a single phase. McRae (8) further stated that when the transport processes in the bulk liquid and vapor spaces are rapid in comparison with those in the membrane, the rate of mass transfer can be expressed as

$$N_A = \frac{1}{L} \int_{C_{Al}}^{C_{Ali}} D \, dc \tag{2.2}$$

where D is the diffusivity of component A in fluid 1 and L is the fictitious film thickness.

If D is assumed to be constant, then equation 2.2 can be simplified as

$$N_A = \frac{D}{L} (C_{Ali} - C_{Al}) \tag{2.3}$$

By setting $k_{cl} = - \frac{D}{L}$, then equation 2.1 and equation 2.3 become the same expression.

At a vapor-liquid surface the resistance to transfer of a component from liquid to vapor is usually taken to be negligible and the vapor and liquid phases are assumed to be in equilibrium.

(b) Transfer of water vapor from the saline solution through the pores to the surface of the condensate.

This step may encounter the following two cases:

(i) If the non-condensable gases are eliminated from the pores of the membrane the water vapor will flow through the pores under the influence of the difference in vapor pressure on both sides of the membrane. Then the mass transfer flux can be expressed as

$$N_A = k_{g1} (P_{A1i} - P_{A2i}) \tag{2.4}$$

where k_{g1} is vapor-phase mass transfer coefficient, P_{A1i} and P_{A2i} refer the interfacial vapor pressure of component A in fluid 1 and 2, respectively.

(ii) If the non-condensable gases are not eliminated from the pores of the membrane, the separation is quite similar to the diffusion through a stagnant gas (9), and the mass transfer flux, N_A, may be determined by

$$N_A = \frac{D_{AB} \pi}{RT Z (P_B)_{1n}} (P_{A1i} - P_{A2i}) \tag{2.5}$$

where π is the total pressure, D_{AB} is the binary gas diffusivity for system A-B, $(P_B)_{1n}$ is the log-mean partial pressure difference of stagnant gas B in the vapor space, R is the gas law constant, Z is the length of diffusing path.

Let

$$k_{g2} = \frac{D_{AB}\,\pi}{RT\ Z\ (P_B)ln} \tag{2.6}$$

Then

$$N_A = k_{g2}\ (P_{A1i} - P_{A2i}) \tag{2.7}$$

Equation 2.4 and equation 2.7 can be expressed in a common equation

$$N_A = k_g\ (P_{A1i} - P_{A2i}) \tag{2.8}$$

(c) Condensation on the pure water side.

Condensation resistance at an interface is assumed negligible as in the case of evaporation resistance at an interface. The transfer of pure water from condensate surface to the condensate bulk must occur by flow, and this resistance is also considered to be negligible.

(d) Overall transfer coefficient.

Sherwood (3) proposed that these resistances in both phases are additive if the flux is constant with time at any small area of the phase boundary. Then the overall mass transfer coefficient, K_m, can be expressed as

$$\frac{1}{K_m} = \frac{1}{k_p 1} + R_m + \frac{1}{k_p 2} \tag{2.9}$$

where boundary layer coefficients kp1 and kp2 are defined as $N_A/(P_{A1} - P_{A1i})$ and $N_A/(P_{A2} - P_{A2i})$, respectively, and R_m is the

resistance of the membrane expressed in terms of the same diffusional potentials as for kp1 and kp2.

Then the mass transfer flux in terms of vapor pressures becomes

$$N_A = K_m (P_{A1} - P_{A2}) \tag{2.10}$$

where P_{A1} is the vapor pressure of the bulk salt solution and P_{A2} is that of the bulk condensate.

The water vapor pressure P_{A1} of salt solution can be expressed as

$$P_{A1} = aP_A \tag{2.11}$$

where a is the activity of water component in salt solution at a given temperature, and P_A is the vapor pressure of pure water at that temperature.

Then equation 2.10 becomes

$$N_A = K_m (aP_A - P_{A2}) \tag{2.12}$$

Equation 2.12 indicates that the factors which should influence permeation rate are:

(1) temperature and pressure, and

(2) thickness and properties of membrane.

Some of these factors have been studied (6, 18), but in this study a typical value of K_m was assumed, and this value was taken as a constant.

C. Heat Transfer Through Porous, Water-Repellent Membranes:

The mechanism of heat transfer through the pores of the membrane is complicated due to the phase changes involved along the path of simultaneous heat and mass transfer. The water vapor diffusing towards the colder surface of the membrane carries latent heat of vaporization in addition to the heat which is normally transferred across the barrier as a result of the prevailing temperature difference. Assuming that radiant energy flux and energy flux of viscous dissipation are of negligible importance, heat is transferred mainly by the following steady state process:

(1) from the warmer fluid (salt solution) to the membrane surface by convection and conduction through the liquid,

(2) through the membrane and pores by conduction,

(3) through the pores, with vapors as latent heat,

(4) from the condensing surface to the colder fresh water by convection and conduction through condensate.

Newton's law of cooling can be applied to processes (1) and (4). Bird (9) and Sherwood (10) have derived an equation for studies of simultaneous heat and mass transfer in multi-component system. Applying the equation to this particular case, the total heat flux can be expressed as

$$q = q_c + q_d \tag{2.13}$$

where q_c is the conductive heat flux across the membrane, and q_d is the heat flux carried by vapor through the pores.

By conduction,

$$q_c = U \Delta T \tag{2.14}$$

where U is the overall heat transfer coefficient and ΔT is the temperature difference between the salt water and fresh water temperatures.

U can be expressed in terms of individual heat transfer coefficients as

$$\frac{1}{U} = \frac{1}{h_1} + \frac{Z}{k} + \frac{1}{h_2} \tag{2.15}$$

where h_1 and h_2 are the film transfer coefficients of fluid 1 and 2, Z is the thickness of the membrane, and k is the effective thermal conductivity of the membrane.

The heat content or enthalpy leaving the solution as a result of mass transfer, q_d, is given by

$$q_d = N_A \Delta H_A \tag{2.16}$$

ΔH_A is the change in the enthalpy of liquid water from inlet to saturated vapor at salt water temperature, and is approximately equal to the latent heat h.

Combining equation 2.12 and equation 2.16,

$$q_d = K_m h(aP_A - P_{A2}) \tag{2.17}$$

Then,

$$q = U\Delta T + K_m h(aP_A - P_{A2}) \qquad (2.18)$$

Equation 2.18 is the usual starting point for engineering studies on heat transfer.

D. Process Economics:

The practical energy requirements and the technological factors governing process and equipment design must, ultimately, be evaluated together on an economic basis to provide an accurate assessment of the potential of this evaporation-through-membranes process for water-desalination applications. This may sometimes be accomplished in a manner similar to that for other mass or energy transport process by dividing the various costs of water production into three major categories, i.e., those which are proportional to the amount of mass or energy transfer surface, those costs which are inversely proportional to the amount of mass or energy transfer surface, and those costs which are independent of the amount of such surface.

When it is possible to divide production costs in the above-described manner, the minimum cost (13) will be obtained when the first two costs are equal. This may be shown mathematically by differentiating the total cost expression in respect to the membrane area and setting the derivative equal to zero. The optimum membrane area required will then be obtained, and its substitution in the total cost expression will give the absolute values of the first two cost components. However, in the method under study,

the driving force is highly non-linear and the costs do not correspond exactly to proportional functions.

Before sound economic analyses can be carried out, the method for cost estimates requires more detailed information on such basic factors as performance of membranes (including permeability of membranes as a function of pressures and temperatures), the time effect on the performance of membrane, stability of membranes at ambient temperatures and elevated temperatures, and the prices of membrane evaporator-condensers as a function of temperation and pressure. Lacking this detailed information at present, desalination costs were calculated from the standardized procedure established by the Office of Saline Water.

The Office of Saline Water (17) has established a procedure for first estimates of desalination cost. This procedure allows the initial estimate on a desalination process to be made with a reasonable expenditure of time and with a minimum amount of engineering data.

Based on this standardized procedure and reasonable assumptions, water desalination costs were evaluated in this study. Although, some of these assumptions require verification on a longer scale, the results using these assumptions indicate the probable feasible ranges of operation and the variables which should be studied to provide more economical operation.

III. METHODS OF STUDY

In this part, two methods are proposed and discussed as a basis for the study of the economics and design conditions of desalination by vaporization through porous, water-repellent membranes. Separate evaporators and heat exchangers are used in Method I, while combined evaporator-heat exchangers are used in Method II. Both methods are based upon the least complex equipment arrangement.

A. Method I:

1. Design of Process:

(a) Process flow:

A flow diagram of this method is shown in Figure 3.1, with typical flows and temperatures.

Fresh sea water enters the pump pits (not shown) through submarine intake lines. A traveling screen filter removes trash, fish and other marine life. A periodic injection of chlorine may be used to prevent growth of marine life in the pit.

Fresh feed is pumped through a preheat section and heated to T_{s2} (100°F in Figure 3.1). The preheated feed then combines with the recycle concentrated sea water and flows through the heat exchanger in which the condensate product flows countercurrently and supplies heat to the combined feed stream. The feed is elevated to T_{s31} (465°F in Figure 3.1) and then through a heater

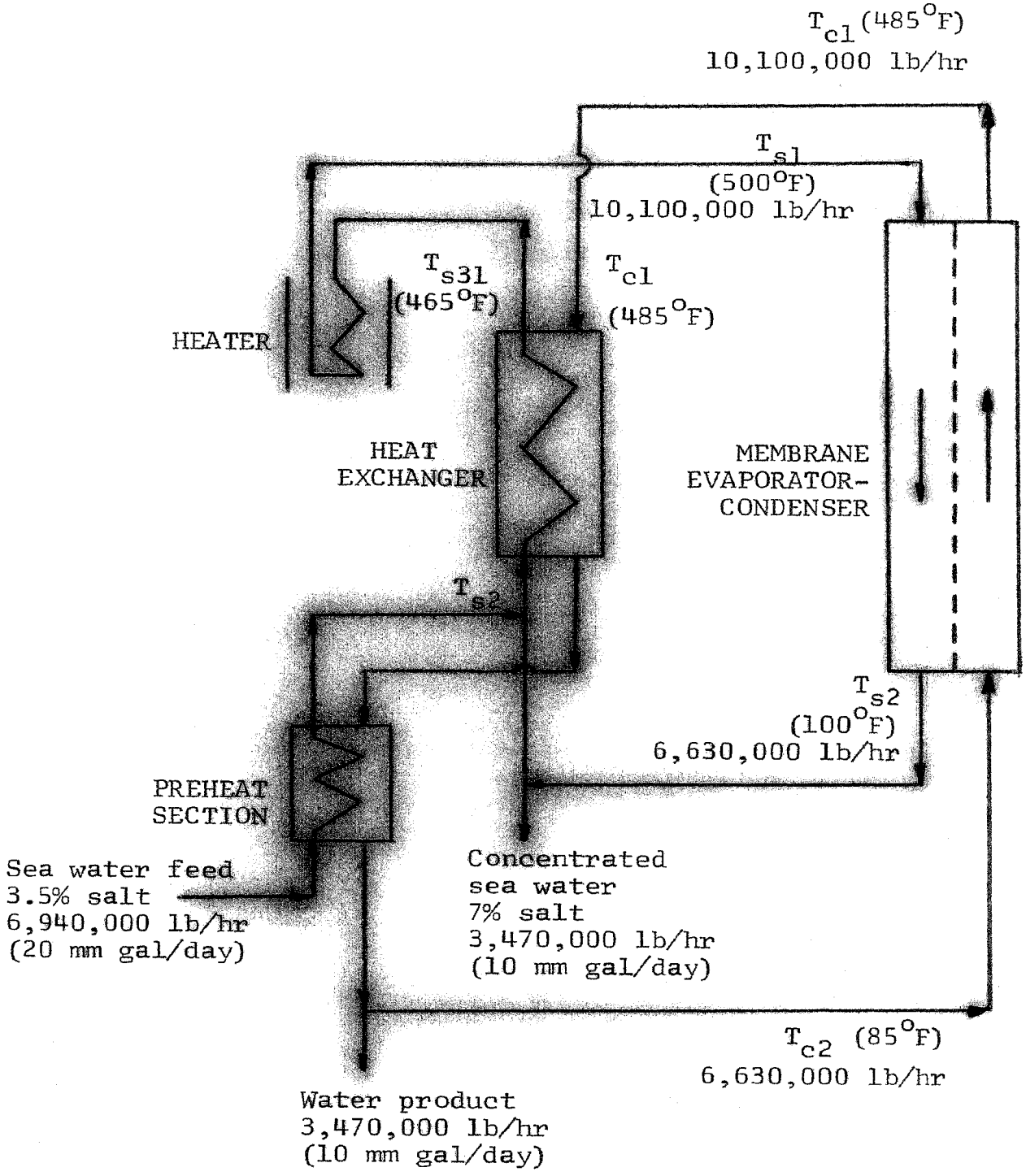


Fig. 3.1 Flow Diagram of Method I

where it is heated to its maximum temperatures, T_{s1} (500°F in Figure 3.1).

The hot feed from the brine heater is distributed to the membrane evaporator-condenser system and flows on one side of the membrane. The water component of the hot saline solution evaporates and passes through the pores of the membrane and condenses on the cold water side. The saline solution, therefore, flows with progressively lower temperature, smaller mass flow and slightly increasing concentration. On the other hand, product water flows countercurrently with progressively higher temperature and mass flow.

It is assumed that concentrated saline solution leaves the evaporator at T_{s2} (100°F in Figure 3.1) and 7% of salt. It is then divided into two streams: One combining with the fresh feed as a recycle and the other leaving the system as concentrated brine.

The water product leaving the evaporator at T_{c1} (490°F in Figure 3.1) first flows through the heat exchanger and then a preheat section where it supplies heat to the sea water feed and is cooled. The water product leaves the preheat section at T_{c2} (85°F in Figure 3.1) and is then divided into two streams. One is recycled to the evaporator as coolant, and the other is pumped to storage and/or usage.

(b) Membrane evaporator-condenser (filter type):

(1) Duty:

A device capable of producing 10 mm gal. per day fresh water by evaporation through porous, water repellent membrane.

(2) Design equation for simultaneous mass and heat transfer:

The simultaneous mass and heat transfer in the membrane evaporator-condenser is shown in Figure 3.2. Hot saline water and cold pure water flow countercurrently with the same mass rate on each side of a porous, water repellent membrane. An exact analysis of this situation has not been well established yet because of the complexity of the desalination process. But the process can be simplified by the following assumptions:

(i) The mass transfer coefficient, K_m, and the heat transfer coefficient, U ,are assumed to be independent of temperature, pressure and flow rate,

(ii) the system is well insulated,

(iii) end effects are of negligible importance,

(iv) the activity of the water in salt solution is independent of temperature and concentration. Consider an infinitesimal section shown in Figure 3.2.

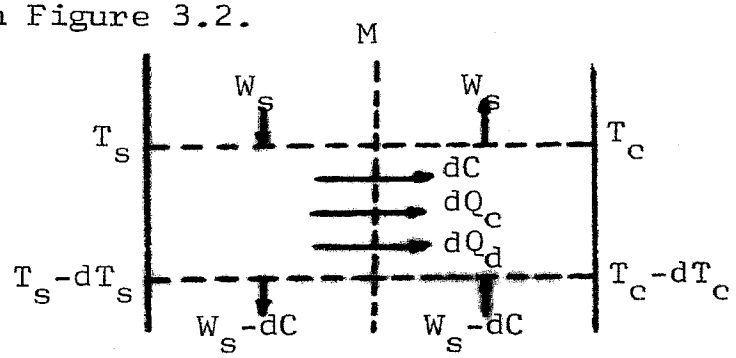


Fig. 3.2 Schematic Diagram for Simultaneous Mass and Heat Transfer of Method I

By applying equation 2.12, the mass transfer is

$$dC = K_m (aP_A - P_{A2}) \, dA \tag{3.1}$$

By applying equation 2.13, the heat transfer by conduction is

$$dQ_c = U (T_s - T_c) \, dA \tag{3.2}$$

By applying equation 2.16, the heat transfer by condensation is

$$dQ_d = K_m h (aP_A - P_{A2}) \, dA \tag{3.3}$$

where Q_c is the rate of heat transfer by conduction, and Q_d is the rate of heat transfer by condensation.

The rate of total heat transfer is the sum of the rate of heat transfer by conduction and that by condensation, assuming that heat transfer by other means is of negligible importance. Then the total heat transfer is

$$\begin{aligned} dQ &= dQ_c + dQ_d \\ &= U (T_s - T_c) dA + K_m h (aP_A - P_{A2}) dA \end{aligned} \tag{3.4}$$

The total heat transfer can also be approximately expressed by a heat balance as

$$dQ = C_p \overline{W}_s dT_s \tag{3.5}$$

where $\overline{W}_s$ is the average salt water flow in an infinitesimal section and C_p is the specific heat of salt water.

From equation 3.4 and equation 3.5,

$$dA = \frac{C_p \overline{W}_s dT_s}{K_m h(aP_A - P_{A2}) + U(T_s - T_c)} \tag{3.6}$$

By substituting equation 3.6 into equation 3.1,

$$dC = \frac{C_p \overline{W}_s dT_s}{h + \dfrac{U(T_s - T_c)}{K_m (aP_A - P_{A2})}} \tag{3.7}$$

where T_s and T_c refer to the saline and the water temperature, respectively.

h is the latent heat of water at T_c.

A is the effective area of porous membrane.

The water vapor pressure and the latent heat expressed as functions of temperature are given in Appendix (B) and (C). The activity of salt water is assumed to be 0.96 as shown in Appendix (A).

ΔC and ΔA can be calculated from equation 2.6 and 2.7 by numerical integration.

Let W_c be the summation of all increments of product water in lb./hr., and A the summation of all increments of membrane area required in $ft.^2$ based on W_{sl} = 1000 lb./hr., the flow rate of salt water entering the evaporator.

The total membrane area required based on a 10 mm gal. per day (3,470,000 lb./hr.) water production basis is then

$$A_m = 3.47 \times 10^6 \times \frac{A}{W_c} \tag{3.8}$$

(c) Auxiliary equipment:

(1) Heat exchanger:

Duty: a device capable of heating fresh feed and concentrated recycle stream from T_{s2} (concentrated recycle temperature leaving the evaporator) to T_{s31} (saline temperature leaving the heat exchanger) by a countercurrent hot product water.

The saline stream leaves the heat exchanger and enters the heater at T_{s31}, and is heated to T_{s1} in the heater. The elevation of temperature, $(T_{s1} - T_{s31})$, in the heater is equal to the sum of the temperature driving forces of D_T and D_{TEX}, where

D_T = the temperature difference in the evaporator-condenser, assumed constant

$$= T_s - T_c$$

D_{TEX} = the temperature difference in the heat exchanger, assumed constant

$$= T_{c1} - T_{s31}$$

Therefore,

$$T_{s31} = T_{s1} - D_T - D_{TEX} \tag{3.9}$$

Let W_{s1} be lb./hr. of fresh feed and concentrated recycle saline. The specific heat of salt water and fresh water in this temperature range is assumed to be 1.1 BTU/lb.,^{o}F. Then, the heat transfer required in BTU/hr. in the heat exchanger

$$= 1.1\ W_{s1}\ (T_{s31} - T_{s2})$$

$$= 1.1\ W_{s1}\ (T_{s1} - D_T - D_{TEX} - T_{s2}) \tag{3.10}$$

The overall heat transfer coefficient

$= 200$ BTU/hr.ft.^{2}^oF.

(estimated from Table 11-10, (16) and p.A-5, (20))

The temperature driving force in heat exchanger

$= D_{TEX}$

The heat exchanger area required in ft.^{2}

$$= \frac{1.1\ W_{s1}\ (T_{s1} - D_T - D_{TEX} - T_{s2})}{200\ D_{TEX}} \qquad (3.11)$$

Let W_c be the product water in lb./hr. based on W_{s1} = 1000 lb./hr. Converted to a 10 mm gal. per day (3,470,000 lb./hr.) water production basis, equation 3.11 becomes

$$A_H = \frac{1.1 \times 1000\ (T_{s1} - D_T - D_{TEX} - T_{s2})}{D_{TEX}} \times \frac{3.47 \times 10^6}{W_c}$$

$$= 1.90 \times 10^7 \times \frac{(T_{s1} - D_T - D_{TEX} - T_{s2})}{D_{TEX}\ W_c} \qquad (3.12)$$

where A_H is the heat exchanger area required in ft.^{2} on a basis of 10 mm gal. per day production.

(2) Heat exchanger --preheat section:

Duty: a device capable of heating 6,940,000 lb./hr. of fresh feed from atmosphere temperature (assume 70^{o}F) to the concentrated recycle temperature, T_{s2}, by countercurrent hot product water.

The average specific heat is assumed to be 1.0 BTU/lb.^{o}F in this low temperature range.

Heat required in BTU/hr.

$$= 1.0 \times 6.94 \times 10^6 \times (T_{s2} - 70)$$

$$= 6.94 \times 10^6 \times (T_{s2} - 70) \qquad (3.13)$$

The overall heat transfer coefficient

$$= 200 \text{ BTU/hr. sq. ft. } ^{o}F$$

(estimated from Table 11-10, (16) and p.A-5, (20))

Temperature driving force = D_{TEX}

Heat exchanger area required in ft.^{2}

$$= \frac{6.94 \times 10^6 \ (T_{s2} - 70)}{200 \times D_{TEX}} \qquad (3.14)$$

$$= 3.47 \times 10^4 \ \frac{(T_{s2} - 70)}{200 \times D_{TEX}}$$

(3) Heater:

Duty: a device capable of elevating fresh feed and concentrated recycle stream, W_{s1}, from T_{s31}, the temperature entering the heater, to T_{s1}, the temperature entering the evaporator. The elevation of temperature in this heater is equal to the sum of the temperature driving forces in the evaporator, D_T, and heat exchanger, D_{TEX}. Therefore,

$$T_{s1} - T_{s31} = D_T + D_{TEX} \qquad (3.15)$$

Specific heat is assumed to be 1.1 in this high temperature range. Heat required

$$= 1.1 \ W_{s1} \ (T_{s1} - T_{s31}) \qquad (3.16)$$

Let W_c be the product water in lb/hr based on W_{s1} = 1000 lb./hr. Converted to 10 mm gal. per day (3,470,000 lb./hr.) water production basis, equation 3.15 becomes

$$Q_H = 1.1 \times 1000 \times (T_{s1} - T_{s31}) \times \frac{3.47 \times 10^6}{W_c}$$

$$= 3.82 \times 10^9 \times (T_{s1} - T_{s31}) / W_c \qquad \text{(a)} \quad (3.17)$$

$$= 3.82 \times 10^9 \times (D_T + D_{TEX}) / W_c \qquad \text{(b)} \quad (3.17)$$

where Q_H is the heat rate in BTU/hr. supplied by heater.

(4) Sea water feed pump:

Duty: a device capable of delivering 20 mm gal./day of sea water at atmospheric pressure to the saturated pressure (680.7 psi) at 500°F.

20 mm gal./day = 13,800 gpm

pressure head = 680.7 - 14.7 = 666 psi

= 666 X 2.31 = 1538 ft. of water

The temperature correction of net positive suction head (NPSH) for centrifugal hot-water pump requires additional 50 ft. of water NPSH. (estimated from p. 6-4, (16)). Assume velocity head is of negligible importance. Total head h_t = 1538 + 50 = 1588 ft. of water.

For 3600 rpm, 3-stage centrifugal pump, specific speed

$$N_s = \frac{rpm\sqrt{gpm}}{(\frac{h_t}{3})^{3/4}} = \frac{3600 \times \sqrt{13,800}}{(\frac{1588}{3})^{3/4}} = 3770$$

(estimated from Figure J-12, (15), efficiency = 80%)

By the inspection of specific speed, 3-stage centrifugal pump operated at 3600 rpm appears desirable and can perform the duty even at this high speed.

Brake horsepower

$$= \frac{\text{gpm X ft. head X sp. gr.}}{3.957 \text{ X efficiency}}$$

$$= \frac{13{,}800 \text{ X } 1588 \text{ X } 1.0}{3.957 \text{ X } 0.81} = 6820 \text{ hp.}$$

(5) Generating engine:

Duty: a device capable of recovering the power consumption by sufficiently utilizing the high pressure exhaust of product water and concentrated sea water. Assume that the generating engine has an efficiency of 75%. Power recovered by either product water or concentrated sea water

$$= \frac{\text{gpm X psi, differential pressure}}{1{,}713} \text{ X efficiency}$$

$$= \frac{10 \text{ X } 10^6}{24 \text{ X } 60} \text{ X } \frac{(680.7 - 14.7)}{1713} \text{ X } 0.75$$

Total power recovered = 2 X 2020 = 4040 hp.

(6) Motor:

Duty: a device capable of driving the sea water feed pump. The net power supplied by motor

$$= 6820 - 4040 = 2780 \text{ hp.}$$

A 300 volt, induction motor operated at 900 rpm can perform this duty. This motor estimated from p. 339, (15) has an efficiency of 90%.

The electrical power consumption

$$= \frac{0.746 \times Hp}{\text{efficiency}} = \frac{0.746 \times 2780}{0.90} = 2300 \text{ Kw.}$$

2. Economic Calculation:

(a) Discussion of variables:

To determine the plant design which will produce water at a minimum cost, many variables affecting the design of the plant were considered. For complete optimization a number of variables, including temperature difference across the membrane, temperature difference in the heat exchanger, inlet and outlet salt water temperatures, mass transfer coefficient, heat transfer coefficient, membrane thickness, salt water flow, thickness and width of channels, and pressure drops, would all have to be considered. However, most of these variables should have only a relatively small effect on the economics of operation and are restricted within certain limits by practical considerations. The most important variables which affect the economics and are free to vary appreciably are the temperatures and temperature differences, and others were assumed constant at values based on previous experiments. The variables which were held constant along with the reason for their restriction are listed below:

(1) Membrane performance:

The performance of a membrane for desalination is mainly controlled by the following effects:

(i) Effect of time:

Since the cost of membranes per unit of time or per unit of production depends upon the life of the membrane over which the cost is prorated, the cost varies widely and could become a very large cost if membrane life were very short. Also, even though the membranes might resist deterioration for a long time, any decrease in performance over a period of time will either decrease the rate of production or require an increase in membrane area. Since no previous studies have been made on the performance of membranes over long periods, an experimental study of long term membrane performance was made and is reported in Section (C), Part III. For economic calculations a membrane life of four months was assumed at an average K_m of 0.3 lb./hr. ft.^{2} in H_g.

(ii) Effect of temperature:

With the same temperature driving force across a membrane, the pressure driving force for desalination will increase as temperature increases. The greater pressure driving force at higher temperatures will produce more fresh water per unit area of membrane, and reduce the costs of the evaporator and membrane for a certain amount of water production, but will involve larger loss of heat to the atmosphere. If the desalination system is assumed to be well insulated, the optimum inlet solution temperature would be the maximum possible. The teflon in the glass fiber membranes would probably deteriorate at 600^{o}F or above so an arbitrary limit of 500^{o}F was assumed. After exposure to sea water at a temperature of 500^{o}F for five days, one membrane was found to be effective and useful for desalination. Three others were tested

for extended periods at 500°F in sea water and were found to maintain their structure and non-wettability. The temperature of salt water entering the membrane evaporator is therefore selected at 500°F for the evaluation of the present process.

Thus far, no study on membrane performance has been made at a temperature higher than 200°F. The reliable relationship between membrane performance and temperature has not yet been well established. It was assumed that membrane performance at elevated temperatures is similar to that below 200°F.

(iii) Effect of flow rate:

Flow rate controlling the flow in either laminar or turbulent region will produce different profiles of velocity, temperature and concentration. The different profiles will give different driving forces and affect the mechanism of mass and heat transfer. Therefore, flow rate can be expected to have some influence on the membrane performance. In a laboratory study, Rohatgi (18) pointed out that the overall mass and heat transfer coefficients increase with flow rate in the laminar flow region. However, the major resistance is in the membrane and flow has only a minor effect. Also these relationships have not been well established on a pilot plant scale. Lacking accurate and sufficient information, the effect of flow rate was assumed to be of negligible importance and the selection of optimum flow rate was omitted in this study. Similar considerations indicate that channel geometry and pressure drop would be of minor importance.

The values of overall mass transfer coefficient used for this study were obtained with negligible flow rates (18), and they should be appropriate for very low flows through parallel evaporators with low pressure drops. The value of overall heat transfer coefficient used was obtained from Rohatgi's results (18) in a laminar flow region.

(iv) Effect of salt concentration:

As discussed before, the mass transfer for desalination is primarily controlled by vapor pressure driving force across the membrane. The vapor pressure on the hot salt water side is a function of salt concentration and temperature and can be determined readily from activities or boiling point elevations.

In the membrane evaporator-condenser, salt concentration increases progressively from 5% to 7% as temperature decreases. In this concentration range, the activity of the salt solution was estimated from p. 11-30, (16) at several temperatures. As shown in Appendix (A), the activity of water remains approximately constant at 0.96.

It should be emphasized here that unknown effects on membrane performance may play very important roles in the evaluation of this desalination process. But, unfortunately, insufficient knowledge at present required the assumption of negligible effects of flow rate and temperature and possibly other factors.

In this study, a conservative value of overall mass transfer coefficient, K_m, and membrane life were obtained from the study

of membrane life as discussed in (C), Part III. The value of overall heat transfer coefficient, U, was obtained from (18). These following values were assumed as the best estimate of performance:

$$K_m = 0.3 \text{ lb./hr. ft.}^2 \text{ in } H_g$$

$$U = 13.0 \text{ BTU/hr. ft.}^2 \ ^oF$$

A conservative membrane life of four months was used. It should be noted that little development work has been done on this type of membrane and improvements seem to be quite probable in the future.

(2) Cost data:

The cost of equipment always depends on size, temperature, pressure, nature of fluid to be handled, and duty to be performed. With the consideration of correction factors of temperature and pressure and corrosion, unit costs were selected from reliable sources and fixed constant. Based on a twenty year plant life, cost data used in this study were given as follows:

Unit cost of evaporator = 2.0 dollars/ft.^{2}

(estimated from (17))

Unit cost of membrane = 5.8 cents per sq. ft. for four months life = 4.73×10^{-4} dollars/ft.^{2} - day

Unit cost of heat exchanger = 2.0 dollars/ft.^{2}

(estimated from p. 11-20, (16))

Unit cost of heater = 1.09 dollars per 1000 BTU/hr. generating capacity

(estimated from p. 319, (19))

Unit cost of heat = 25.0 cents per 1,000,000 BTU

(estimated from (17))

Unit cost of continuous demand of electric power below 100,000 K_w is 0.7 cents/K_w-hr.

(estimated from (17))

The costs of power, pumps, generating engine, and motor were considered to be of little influence on water cost and were held constant.

(b) Cost equation:

Based on a twenty year plant life, the necessary items of cost for a 10 mm gallons per day desalting plant are given below for the estimation of water cost.

Cost of evaporator, C_{OEV}, in dollars is obtained by the product of total membrane area required, A_m, given in equation 3.8 and the unit cost of evaporator. Therefore,

$$C_{OEV} = 3.47 \times 10^6 \times \frac{A}{W_c} \times 2.0$$

$$= 6.94 \times 10^6 \times \frac{A}{W_c} \qquad (3.18)$$

Similarly, the cost of membrane, C_{ME}, in dollars/day can be obtained by applying equation 3.8 and the unit cost.

Cost of membrane

$$C_{MB} = 3.47 \times 10^6 \times \frac{A}{W_c} \times \frac{0.058}{120}$$

$$= 1675 \frac{A}{W_c} \qquad (3.19)$$

The cost of main heat exchanger, C_{OE1}, in dollars can be obtained by applying equation 3.12 and the unit cost of heat exchanger.

$$C_{OE1} = 1.90 \times 10^7 \frac{(T_{s1} - D_T - D_{TEX} - T_{s2})}{D_{TEX} W_c} \times 2.0$$

$$= 3.80 \times 10^7 \frac{(T_{s1} - D_T - D_{TEX} - T_{s2})}{D_{TEX} W_c} \qquad (3.20)$$

The cost of preheat exchanger, C_{OE3}, in dollars can be obtained by applying equation 3.14 and the unit cost of heat exchanger.

$$C_{OE3} = 3.47 \times 10^4 \times \frac{(T_{s2} - 70)}{D_{TEX}} \times 2.0$$

$$= 6.94 \times 10^4 \times \frac{(T_{s2} - 70)}{D_{TEX}} \qquad (3.21)$$

Cost of heater, C_{OHR}, in dollars can be obtained by applying equation 3.17 and the unit cost of heater.

$$C_{OHR} = 3.82 \times 10^9 \times \frac{(D_T + D_{TEX})}{W_c} \times \frac{1.09}{1000}$$

$$= 4.16 \times 10^6 \times \frac{(D_T + D_{TEX})}{W_c} \qquad (3.22)$$

Cost of heat, C_{HT}, in dollars per day can be obtained by applying equation 3.17 and the unit cost of heat.

$$C_{HT} = 3.82 \times 10^9 \times \frac{(D_T + D_{TEX})}{W_c} \times \frac{0.25}{10^6} \times 24$$

$$= 2.29 \times 10^4 \times \frac{(D_T + D_{TEX})}{W_c} \qquad (3.23)$$

Sea water/feed pump, generating engines and motor that can perform the duty described before have the following cost.

$$C_{OPM} = 1.50 \times 10^5 \qquad (3.24)$$

$$C_{OEG} = 1.20 \times 10^5 \qquad (3.25)$$

$$C_{OMO} = 3.90 \times 10^4 \qquad (3.26)$$

where C_{OPM} is the cost of pump in dollars,

C_{OEG} is the cost of engine in dollars, and

C_{OMO} is the cost of motor in dollars.

All these items of cost are estimated from (17).

The electrical power consumption of motor is 2300 Kw. The cost of power, C_{PO}, in dollars per day is therefore,

$$C_{PO} = 2300 \times \frac{0.7}{100} \times 24 = 387 \qquad (3.27)$$

Following the procedure given by the Office of Saline Water (17), water cost can be evaluated by the following equations.

Principal items of equipment,

$$P_{IE} = C_{OEV} + C_{OE1} + C_{OE3} + C_{OHR} + C_{OPM} + C_{OMO} + C_{OEG} \qquad (3.28)$$

Total operating costs for one stream day,

$$T_{OP} = C_{PO} + C_{MB} + C_{HT} + 5.90 \times 10^{-4} \times P_{IE} + 380.2 \quad (3.29)$$

Cost of product water per 1,000 gallons,

$$C_{MG} = T_{OP} \times 10^{-4} \quad (3.30)$$

(c) Cost optimization:

Substitute equation 3.18 through 3.29 into equation 3.30, cost of product water per 1,000 gallons can be expressed as:

$$C_{MG} = 0.0949 + 0.576 \frac{A}{W_c} + 2.54 \frac{(D_T + D_{TEX})}{W_c} + 2.24 \frac{(T_{s1} - D_T - D_{TEX} - T_{s2})}{W_c D_{TEX}} + 0.00408 \frac{(T_{s2} - 70)}{D_{TEX}} \quad (3.31)$$

In equation 3.31, A and W_c are functions of T_{s1}, T_{s2} and D_T, and can be obtained from equation 3.6 and equation 3.7 by numerical integration. If T_{s1} is specified, the variables remaining for optimization are:

(1) the temperature driving force, D_T, in the membrane evaporator-condenser,

(2) the temperature difference, D_{TEX}, in the heat exchanger,

(3) brine temperature, T_{s2}, leaving the membrane evaporator-condenser.

The assumed feasible region for searching optimum point is

$$5 \leq D_T \leq 25^{\circ}F$$

$$10 \leq D_{TEX} \leq 40^{\circ}F$$

$$80 \leq T_{s2} \leq 500^{\circ}F$$

A computer program was written which would calculate costs under various levels of these three variables. Given the necessary data, the computer computes the heat and mass balance, determines the required membrane and heat exchanger areas, and computes the cost of the water produced by this plant. It then changes the variables D_T and D_{TEX}, and recalculates the requirements. This was accomplished in the following steps:

(i) Set initial value of the three variables (D_T, D_{TEX}, T_{s2}).

(ii) Calculate the membrane evaporator-condenser design.

(iii) Incrementally with evaporator outlet solution temperature calculate all the necessary items for evaluating the water cost.

(iv) Calculate the water cost at various levels of evaporator outlet solution temperature.

(v) Increase D_{TEX} by $10^{\circ}F$.

(vi) Repeat from step (i) up to $D_{TEX} = 40^{\circ}F$.

(vii) Increase D_T by $5^{\circ}F$.

(viii) Repeat from step (i) up to $D_T = 25^{\circ}F$.

(ix) Search the domain of optimum conditions.

(x) Search the true optimum conditions by using small

increments of $1^{o}F$ in the domain of optimum conditions.

B. Method II:

1. Design of Process:

(a) Process flow:

A flow diagram of this method is shown in Figure 3.3.

The pretreatment of the fresh feed in this method is same as that in Method I. After pretreatment, feed sea water is pumped through a preheat-exchanger in which the water product and the concentrated sea water flow countercurrently and supply heat to the feed. The feed is elevated to the appropriate inlet temperature.

The preheated feed then flows through the heat exchanger tubes in the evaporator-condenser system in which it picks up heat from the product water, as it flows countercurrently toabbth product water and evaporating solution. This maintains the temperature of the product water sufficiently low for condensation. The feed leaves the evaporator-condenser system at near the maximum temperature and enters a heater where it is heated to maximum temperature.

The hot feed from the heater is fed back into the center portion of the evaporator-condenser. The water component then evaporates and passes through the pores of the membrane and condenses on the cold water side. The evaporating saline solution, therefore, flows with progressively lower temperatures, smaller flow rates and higher salt concentration from pore to pore. The

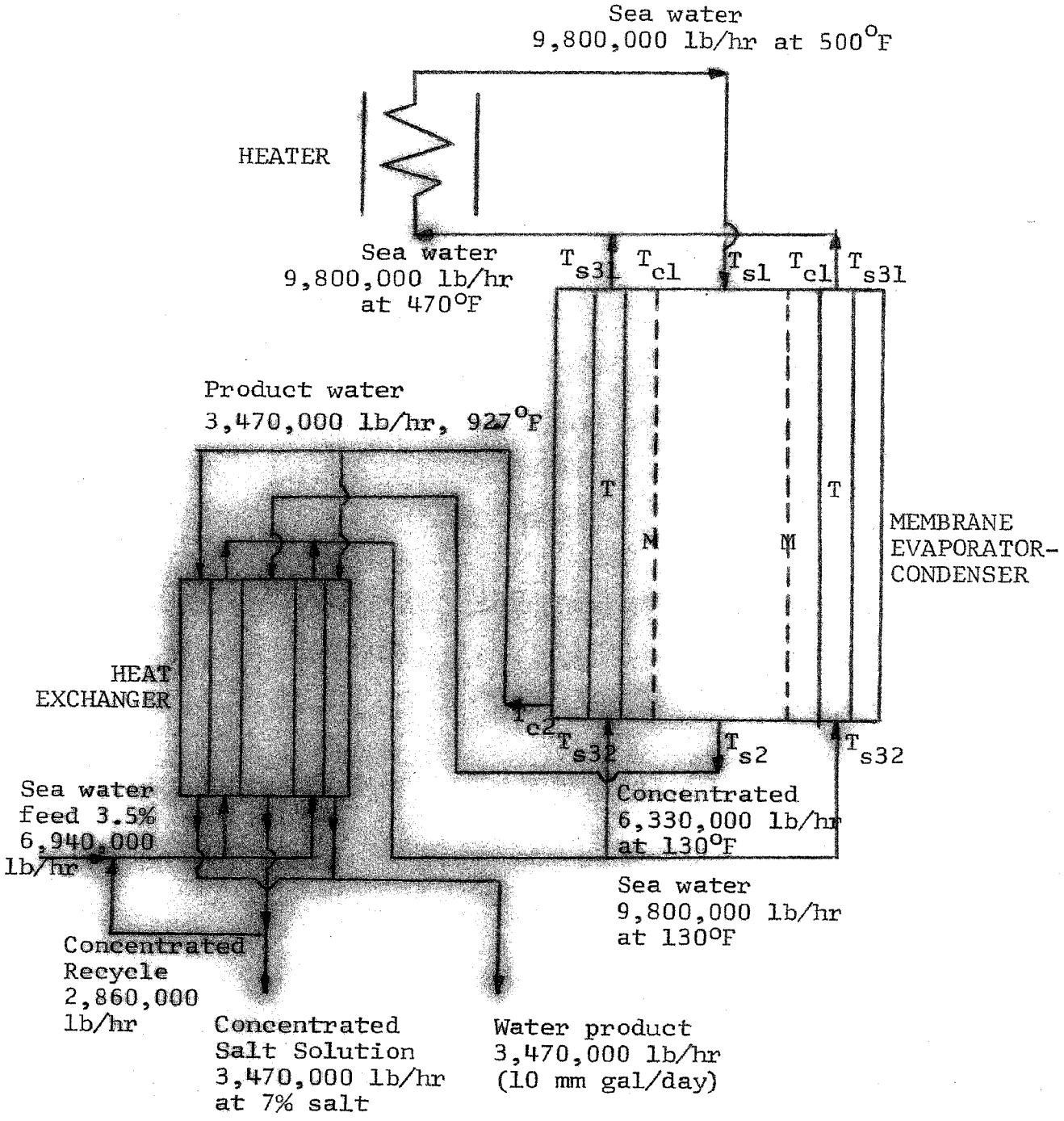


Fig. 3.3 Flow Diagram of Method II

concentrated sea water leaves the evaporator-condenser at a low temperature and with a salt concentration less than 7%. However, a 7% salt solution was assumed for calculations.

In the evaporator-condenser, the water product stream absorbs heat of condensation and conduction from the hot concurrent saline solution inside the membrane. The product water transfers the absorbed heat plus its own sensible heat to the countercurrent feed in the heat exchanger tubes. In this manner, the product stream flows with progressively lower temperature, and higher mass flow, and leaves at low temperature.

Both concentrated sea water and product water then flow through the preheat exchanger. The concentrated sea water may then be divided into two streams, one being recycled while the other with the separate product water passes through reversed positive displacement pumps acting as engines to recover work as the pressure is reduced. The concentrated sea water leaves the desalination system as a concentrated product. The product water is pumped to storage and/or usage.

(b) Membrane evaporator-condenser (cylinder type):

(1) Duty: a device capable of producing 10 mm gal/day fresh water by evaporation through porous, water-repellent membrane.

(2) Design equation for membrane evaporator-condenser:

A sketch which is useful in describing the simultaneous mass and heat transfer process occuring in the membrane evaporator-condenser is shown in Figure 3.4.

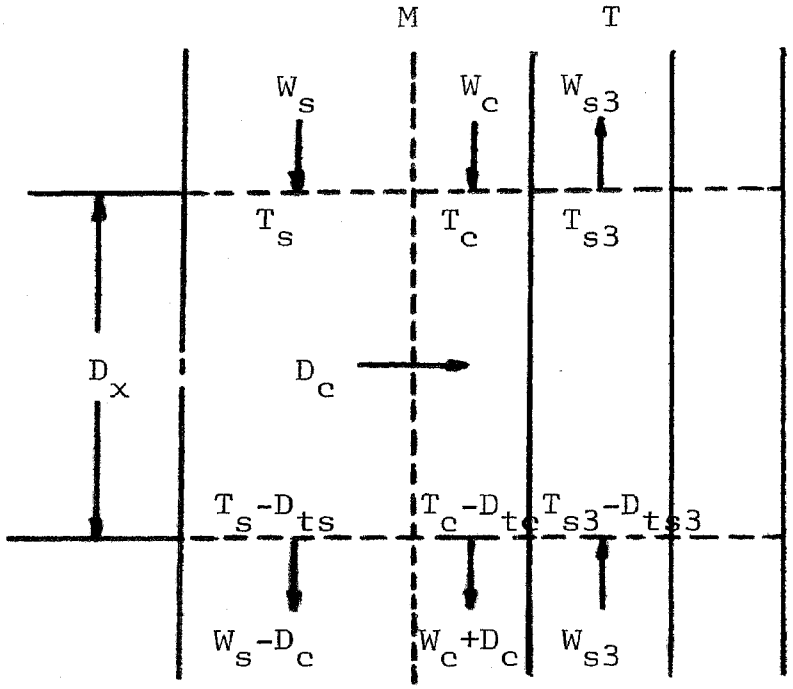


Fig. 3.4 Schematic Diagram for Simultaneous Mass and Heat Transfer of Method II

The three streams are described separately below.

(i) Hot salt water, W_s, inside the membrane M flows with progressively lower temperatures (by the loss of heat as a result of evaporation and conduction) and lower mass flow by the evaporation of water component.

(ii) The concurrent product water, W_c, on the shell side absorbs the condensate and flows with increasing mass flow. But it donates not only the absorbed heat but also its own sensible heat to the fresh salt water, W_{s3}, in the heat exchanger tubes T. Therefore, it flows with decreasing temperature.

(iii) The countercurrent fresh salt water, W_{s3}, in the heat exchanger tubes, T, flows with constant mass flow but progressively higher temperatures in the opposite direction to evaporating solution.

The first three assumptions in Method I were maintained in this method. A boiling-point-elevation correction equation was used instead of the fourth assumption of constant activity. An additional important assumption is that the membrane performance in this concurrent salt water and product water system is the same as assumed for the countercurrent system described in Method I and the effect of curvature of the membrane tube is assumed of negligible importance. Differential equations for this system are shown in Appendix G. However, for convenience, the analysis and calculations were carried out incrementally, using finite difference equations.

Let D_c be the water product transferred in a small incremental section of D_x.

By an overall heat balance for all three streams, the following equation is obtained by taking zero reference temperature and assuming that specific heats of salt water and fresh water are equal and constant.

$$C_p W_s T_s - C_p (W_s - D_c)(T_s - D_{ts}) + C_p W_c T_c - C_p (W_c + D_c)(T_c - D_{tc}) - C_p W_{s3} D_{ts3} = 0 \qquad (3.32)$$

where W is flow in lb/hr.,

T is temperature in ^{o}F,

C_p is specific heat of water or solution, in BTU/lb., ^{o}F,

D_t is change in stream temperature.

Subscripts s, c, and s3 refer to evaporating solution, condensate, and solution being heated within the increment.

The elevating temperature, D_{ts3}, of salt water being heated, W_{s3}, is achieved by heat transferred from the water product side through the tubes having an overall heat transfer coefficient, U_E, and a total perimeter P_E. A mean temperature difference, $\left[(T_c - 0.5\ D_{tc}) - (T_{s3} - 0.5\ D_{ts3})\right]$, serves as the driving force for heat transfer.

Thus,

$$C_p W_{s3} D_{ts3} = U_E P_E D_x \left[(T_c - 0.5\ D_{tc}) - (T_{s3} - 0.5\ D_{ts3}) \right] \qquad (3.33)$$

By applying equation 2.10, water product in this small section, D_x, is

$$D_c = K_m P_m D_x \ (P_{A1} - P_{A2}) \tag{3.34}$$

where P_m is the perimeter of membrane tube.

$P_{A1} = (P)_{T_s - 0.5\ D_{ts} - E}$, the vapor pressure of water at the average salt water temperature minus the boiling point elevation of salt water, E, and

$P_{A2} = (P)_{T_c - 0.5\ D_{tc}}$, the vapor pressure of water at the average fresh water temperature.

In a small section, pressure difference is assumed to be linear with the temperature difference. Then for water, $\Delta P_{T_a - T_b} \cong (T_a - T_b) \frac{\partial P}{\partial T}$, where $\frac{\partial P}{\partial T} \cong \frac{P_1 - P_2}{T_1 - T_2}$ with T_1 and T_2 in the region of T_a and T_b. Therefore,

$$P_{A1} - P_{A2} = \left[(T_s - 0.5\ D_{ts} - E) - (T_c - 0.5\ D_{tc})\right] \times \frac{P_1 - P_2}{T_1 - T_2}$$

$$= \frac{(T_s - 0.5\ D_{ts} - E) - (T_c - 0.5\ D_{tc})}{(T_s - 0.5\ D_{ts}) - (T_c - 0.5\ D_{tc})} \ (P_1 - P_2) \tag{3.35}$$

where P_1 and P_2 are water vapor pressures at temperatures $T_1 = (T_s - 0.5\ D_{ts})$ and $T_2 = (T_c - 0.5\ D_{tc})$, respectively.

Assume that D_{ts} and D_{tc} are approximately equal in the denominator of equation 3.35 in order to avoid higher order equations. Then,

$$P_{A1} - P_{A2} = \frac{P_1 - P_2}{T_s - T_c} \left[(T_s - 0.5\ D_{ts} - E) - (T_c - 0.5\ D_{tc}) \right] \tag{3.36}$$

By substituting equation 3.36 into equation 3.34,

$$D_c = K_m P_m P_x \frac{P_1 - P_2}{T_s - T_c} \left[(T_s - 0.5\ D_{ts} - E) - (T_c - 0.5\ D_{tc}) \right] \tag{3.37}$$

Let,

$$C_k = K_m P_m \frac{P_1 - P_2}{T_s - T_c} \tag{3.38}$$

Then,

$$D_c = C_k D_x \left[(T_s - 0.5\ D_{ts} - E) - (T_c - 0.5\ D_{tc}) \right] \tag{3.39}$$

where C_k is defined as an overall mass transfer coefficient expressed in lb./hr.^{o}F. ft.

Consider the salt stream inside the membrane as a system. The heat of evaporation and conduction of the system is supplied by sensible heat. A heat balance over the system can be expressed as

$$C_p W_s T_s - C_p (W_s - D_c)(T_s - D_{ts}) = D_c \left[h + (T_s - 0.5\ D_{ts})\ C_p \right] + U P_m D_x \left[(T_s - 0.5\ D_{ts}) \right.$$

$$- (T_c - 0.5\ D_{tc})\Big] \tag{3.40}$$

where $h + (T_s - 0.5\ D_{ts})\ C_p$ is the approximate enthalpy of water evaporated per lb. with the latent heat, h, at the mean temperature, $(T_s - 0.5\ D_{ts})$, of the system. Therefore, the total heat loss by evaporation is $D_c\left[h + (T_s - 0.5\ D_{ts})\ C_p\right]$. The heat loss by conduction is $UP_m D_x\left[(T_s - 0.5\ D_{ts}) - (T_c - 0.5\ D_{tc})\right]$ with mean temperatures, $(T_s - 0.5\ D_{ts})$ on the salt water side and $(T_c - 0.5\ D_{tc})$ on the product water side.

The four variables, D_c, D_x, D_{tc} and D_{ts3}, in equations 3.32, 3.33, 3.39, and 3.40 are unknown and are to be solved for each increment. In order to ensure that the solutions converge, an approximate method is first used to estimate D_c as follows:

Rewrite equation 3.36 by setting $D_{ts} = D_{tc}$, then

$$D_c = C_k D_x\ (T_s - T_c - E) \tag{3.41}$$

$$D_x = \frac{D_c}{C_k\ (T_s - T_c - E)} \tag{3.42}$$

Then substitute equation 3.42 into equation 3.40, solve for D_c, and let $D_{tc} = D_{ts}$, to give

$$D_c = \frac{W_s D_{ts} C_p}{h + 0.5\ D_{ts} C_p + \dfrac{UP_m(T_s - T_c)}{C_k(T_s - T_c - E)}} \tag{3.43}$$

Then equation 3.42 can be used as an estimate of D_x.

Substitute equation 3.33 and equation 3.40 into equation 3.32.

$$D_c \left[h + (T_s - 0.5\ D_{ts})\ C_p\right] + UP_m D_x \left[(T_s - 0.5\ D_{ts}) - (T_c - 0.5\ D_{tc})\right] + C_p W_c T_c - C_p\ (W_c + D_c)(T_c - D_{tc}) - U_E P_E D_x \left[(T_c - 0.5\ D_{tc}) - (T_{s3} - 0.5\ D_{ts3})\right] = 0 \qquad (3.44)$$

On simplification,

$$D_c \left[h + (T_s - 0.5\ D_{ts} - T_c + D_{tc})\ C_p\right] + UP_m D_x \left[(T_s - 0.5\ D_{ts}) - (T_c - 0.5\ D_{tc})\right] + C_p W_c D_{tc} - U_E P_E D_x \left[(T_c - 0.5\ D_{tc}) - (T_{s3} - 0.5\ D_{ts3})\right] = 0 \qquad (3.45)$$

Assume that $D_{tc} = D_{ts}$ in the first term and $D_{ts3} = D_{ts}$ in the last term in equation 3.45. Then substitute equation 3.41 into equation 3.45.

$$C_k D_x \left[(T_s - 0.5\ D_{ts} - E) - (T_c - 0.5\ D_{tc})\right] x \left[h + (T_s - T_c + 0.5\ D_{ts})\ C_p\right] + UP_m D_x \left[(T_s - 0.5\ D_{ts}) - (T_c - 0.5\ D_{tc})\right] + C_p W_c D_{tc} - U_E P_E D_x \left[(T_c - 0.5\ D_{tc}) - (T_{s3} - 0.5\ D_{ts})\right] = 0 \qquad (3.46)$$

Solve for D_{tc} from equation 3.46.

$$D_{tc} = \Big\{ U_E P_E D_x \ (T_c - T_{s3} + 0.5\ D_{ts}) - C_k D_x \Big[h + (T_s + 0.5\ D_{ts} - T_c)\ C_p \Big] (T_s - T_c - 0.5\ D_{ts} - E) - UP_m D_x \ (T_s - T_c - 0.5\ D_{ts}) \Big\} \Big/ \Big\{ 0.5 \Big[h + (T_s - T_c + 0.5\ D_{ts})\ C_p \Big] C_k D_x + 0.5\ (UP_m + U_E P_E) D_x + C_p W_c \Big\} \qquad (3.47)$$

After D_c, D_x and D_{tc} have been estimated, D_x can be estimated more accurately by equation 3.39, and D_{ts3} can be solved from equation 3.32.

$$D_{ts3} = \frac{W_s D_{ts} + D_c\ (T_s - D_{ts} - T_c + D_{tc}) + W_c D_{tc}}{W_{s3}} \qquad (3.48)$$

The original estimates for D_c and D_{tc} were found to be sufficiently accurate for design and convergence purposes.

An equilibrium temperature, T_c, corresponding to T_s and T_{s3} at the starting point was found by using the following heat balance equation, which assumes a steady state T_c with no flow, W_c.

$$C_k D_x \ (T_s - T_c - E)\ h + UP_m D_x \ (T_s - T_c) - U_E P_E D_x \ (T_c - T_{s3}) = 0 \qquad (3.49)$$

The first term in quation 3.49 is the heat of condensation liberated by water vapor from the hot salt water. The second term is the heat of conduction established by the temperature driving

force, $(T_s - T_c)$, across the membrane. And the third term is the heat transferred from product water to fresh salt water in the heat exchanger tubes.

Solve for T_c from equation 3.49.

$$T_c = \frac{U_E P_E T_{s3} + C_k (T_s - E) h + UP_m T_s}{h C_k + UP_m + U_E P_E} \tag{3.50}$$

The boiling point elevation, E, of 7% sea water at any temperature T_s can be estimated from the following equation (23).

$$E = 1.0 + 0.00267 T_s \tag{3.51}$$

The temperatures and the water product were accomplished by the following steps:

(1) Select a starting value of T_s and T_{s3}.

(2) Calculate the equilibrium temperature, T_c, from equation 3.50 as a starting point for product water stream.

(3) For a given constant increment D_{ts}, calculate D_c from equation 3.43.

(4) Calculate D_x from equation 3.42.

(5) Calculate D_{tc} from equation 3.47.

(6) Calculate D_{ts3} from equation 3.48.

(7) Recalculate D_x from equation 3.39.

(8) Calculate new starting values of T_s, T_c and T_{s3}.

(9) Record results at specified temperatures and repeat from step (3).

Several calculation procedures were used, most of which were unstable and produced impossible values of T_c. According to manual estimates of conditions at various points, the above procedure seemed the most reliable. For values of increment size much larger than $D_{ts} = 2^{o}F$, this procedure may become unstable.

Let W_c be the summation of all increments of water product, D_c, in lb./hr. and X the summation of all increments of evaporator length, D_x, in ft. based on $W_{sl} = 1000$ lb./hr., the flow rate of salt water entering the evaporator. Converted to 10 mm gallons per day (3,470,000 lb./hr.) water production basis,

Total evaporator length required

$$X_T = 3.47 \times 10^6 \frac{X}{W_c} \tag{3.52}$$

Total membrane area required

$$A_m = P_m X_T \tag{3.53}$$

Total heat exchanger area required

$$A_{hm} = P_E X_T \tag{3.54}$$

(c) Auxiliary equipment:

In this method, the evaporator serves another important function similar to that of a heat exchanger to elevate the temperature of fresh sea water. This duty can be performed by the tubes, on the shell side of the evaporator, through which heat is transferred from hot product water to cold fresh sea water in the tubes. The main heat exchanger is thus eliminated.

Heater, pump, engine and motor were assumed to have the same duty as that discussed in Method I. In the evaluation of the heat rate supplied by heater, only equation 3.17 (a) can be applied in this case. The power consumption of the motor is, of course, equal to that of Method I due to the assumed same duty of the motor. Only the preheat exchanger has a slightly different duty and this is discussed below.

As shown in Figure 3.4, hot concentrated recycle combines with the feed sea water (assumed at 70^{o}F). The combined stream is assumed to be 75^{o}F and then enters the preheat exchanger. In the preheat exchanger, this stream is heated to T_{s32}, the saline temperature entering the tubes of the evaporator, by separate streams of hot concentrated exhaust and product water.

Let W_{s1} be the flow in lb./hr. of the combined fresh feed and concentrated recycle stream. Its average specific heat is assumed to be 1.1 BTU/lb.,^{o}F. Then, the heat required to elevate W_{s1} from 75^{o}F to T_{s32} is 1.1 W_{s1} $(T_{s32} - 75)$ BTU/hr. The overall heat transfer coefficient is 200 BTU/hr.,sq. ft.,^{o}F. (estimated from Table 11-10, (16) and p. A-5, (20)).

Assume that the temperature difference for heat exchange between the average temperature of concentrated exhaust and product water and the temperature of feed sea water at the preheater exit is sufficiently accurate for preheater calculations. Then, the average temperature of concentrated exhaust and product water is equal to $(W_{s2}\ T_{s2} + W_{c}\ T_{c2})/(W_{s2} + W_{c})$, where T_{s2} and T_{c2}

represent the temperature of concentrated exhaust, W_{s2}, and that of product water, W_c, leaving the evaporator, respectively. Therefore, temperature difference for heat exchanger is equal to $((W_{s2}\ T_{s2} + W_c\ T_{c2})/\ (W_{s2} + W_c) - T_{s32})$, the heat exchanger area required in ft.^{2} is equal to

$$\frac{1.1\ W_{s1}\ (T_{s32} - 75)}{200\ \left(\frac{W_{s2}\ T_{s2} + W_c\ T_{c2}}{W_{s2} + W_c} - T_{s32}\right)}$$

And

$$W_{s1} = W_{s3} + W_c \tag{3.55}$$

Let W_c be the product water flow rate in lb./hr. based on W_{s1} = 1000 lb./hr. Converted to a 10 mm gallons per day (3,470,000 lb./hr.) water production basis,

$$A_H = \frac{1.1 \times 1000\ (T_{s32} - 75)}{200\ \left(\frac{W_{s2}\ T_{s2} + W\ T_{c2}}{1000} - T_{s32}\right)} \times \frac{3.47 \times 10^6}{W_c}$$

$$= 1.90 \times 10^7\ \frac{(T_{s32} - 75)}{W_c\ \left(\frac{W_{s2}\ T_{s2} + W_c\ T_{c2}}{1000} - T_{s32}\right)} \tag{3.56}$$

where A_H is the heat exchanger area required in ft.^{2} on a basis of 10 mm gallons per day water production.

2. Economic Calculation:

(a) Cost equation:

All the cost data are listed under Method I. Additional data which is necessary was estimated from reference (21) and is given below:

Unit cost of evaporator = 2.0 dollars per ft. for 4 inches iron pipe plus fittings.

Unit cost of heat exchanger tube = 0.170 dollars per ft. for 3/4 in. Titanium tube.

0.115 dollars per ft. for 1/2 in. Titanium tube.

On a basis of 1000 lb./hr. sea water entering the evaporator at 500°F, the water product, W_c, and the required length of evaporator, X, can be obtained as described in the previous section. These values are converted to a 10 mm gallons per day production basis. All the main items of cost can be calculated at different sea water temperatures, T_{s2}, leaving the evaporator, for given values of perimeter of membrane tube, P_m, perimeter of heat exchanger tubes, P_E, and T_{s31}, the sea water temperature at the evaporator tube exit. The costs are obtained by multiplying the requirements for desalination with the unit costs.

By applying equation 3.52, the cost of evaporator, C_{OEV}, is obtained as follows:

$$C_{OEV} = (2.0 + C_t N)\ X_T$$

$$= \frac{(2.0 + C_t N)\ X}{W_c} \times 3.47 \times 10^6 \qquad (3.57)$$

where C_t is unit cost, N heat exchanger tubes.

By applying equation 3.53, the cost of membrane, in dollars/day, is

$$C_{mb} = 4.73 \times 10^{-6}\ A_m$$

$$= 4.73 \times 10^{-4}\ P_m\ X_T$$

$$= 4.73 \times 10^{-4}\ P_m \times 3.47 \times 10^6\ \frac{X}{W_c}$$

$$\doteq 1675\ P_m\ \frac{X}{W_c} \qquad (3.58)$$

By applying equation 3.56, the cost of preheat exchanger is

$$C_{OEX} = 2.0\ A_h$$

$$= 2.0 \times 1.90 \times 10^7\ \frac{(T_{s32} - 75)}{W_c\left(\frac{W_{s2}\ T_{s2} + W_{c2}\ T_{c2}}{1000} - T_{s32}\right)}$$

$$= 3.80 \times 10^7\ \frac{(T_{s32} - 75)}{W_c\left(\frac{W_{s2}\ T_{s2} + W_{c2}\ T_{c2}}{1000} - T_{s32}\right)} \qquad (3.59)$$

By applying equation 3.17 (a), the cost of heater is

$$C_{OHR} = \frac{1.09}{1000}\ Q_h$$

$$= \frac{1.09}{1000} \times 3.82 \times 10^9 \frac{(T_{s1} - T_{s31})}{W_c}$$

$$= 4.16 \times 10^6 \frac{(T_{s1} - T_{s31})}{W_c} \qquad (3.60)$$

The cost of heat in dollars per day is

$$C_{HT} = \frac{0.25}{1000} \times Q_h \times 24$$

$$= \frac{0.25}{1000} \times 3.82 \times 10^9 \frac{(T_{s1} - T_{s31})}{W_c} \times 24$$

$$= 2.29 \times 10^4 \frac{(T_{s1} - T_{s31})}{W_c} \qquad (3.61)$$

Costs of pump, engine, motor and power consumption were assumed to be the same as that of Method I. Therefore, equations 3.24 to equation 3.27 can be applied to this method.

Principal items of equipment,

$$P_{IE} = C_{OEV} + C_{OEX} + C_{OHR} + C_{OPM} + C_{OMO} + C_{OEG} \qquad (3.62)$$

Water cost can then be calculated from equation 3.29 and equation 3.30.

Thus, in the next section water cost will be minimized with respect to the pertinent variables.

(b) Cost optimization:

The membrane performance and cost data which are considered

fixed for seeking the optimum cost have been discussed in Method I. An additional variable, the diameter of evaporator, was assumed to have little effect on the product water cost. Four inch iron pipe was selected as an evaporator, since fresh water is in contact with the outer pipe.

Substitute equation 3.24 through equation 3.27 and equation 3.57 through equation 3.62 into equation 3.29 and equation 3.30, cost of product water per 1000 gallons can be rewritten as:

$$C_{mg} = 0.0949 + 0.1675 \frac{P_m X}{W_c} + 2.54 \frac{(T_{s1} - T_{s31})}{W_c}$$

$$+ 0.205 \frac{(2.0 + C_t N)\ X}{W_c}$$

$$+ 2.24 \frac{(T_{s32} - 75)}{W_c \left(\frac{W_{s2}\ T_{s2} + W_c\ T_{c2}}{1000} - T_{s32}\right)} \qquad (3.63)$$

In equation 3.63, X, W_c, T_{c2} and T_{s32} are functions of T_{s1}, T_{s2}, T_{s31}, P_m and P_E, and can be obtained from equations 3.32 to 3.51 by the method discussed in the Design Equation of Method II. After T_{s1} has been specified, the variables remaining for optimization are:

(i) salt water temperature, T_{s2}, leaving the evaporator,

(ii) heated salt water temperature, T_{s31}, leaving the heat exchanger tubes in the evaporator,

(iii) perimeter P_m of membrane tube,

(iv) perimeter P_E of heat exchanger tubes.

The assumed feasible region for searching optimum point is

$$80 \leq T_{s2} \leq 500^{o}F$$

$$440 \leq T_{s31} \leq 490^{o}F$$

$$0.523 \leq P_{m} \leq 1.088 \text{ ft.}$$

$$0.785 \leq P_{E} \leq 2.093 \text{ ft.}$$

A computer program was written for obtaining the optimum cost. A "simultaneous" direct search technique was used, i.e., the cost at various points over the entire feasible region was calculated and the minimum cost determined by directly examining the cost calculation output. This was accomplished in the following steps:

(i) Set initial value of the four variables.

(ii) Calculate the membrane evaporator length incrementally with $D_{ts} = 2^{o}F$.

(iii) Calculate all the necessary items for evaluating the water cost at each $50^{o}F$ increment.

(iv) Calculate the water cost at each $50^{o}F$ increment.

(v) Change one of the variables.

(vi) Repeat from step (i), record results at specified intervals of temperature.

These results were used to visually select an optimum cost and design.

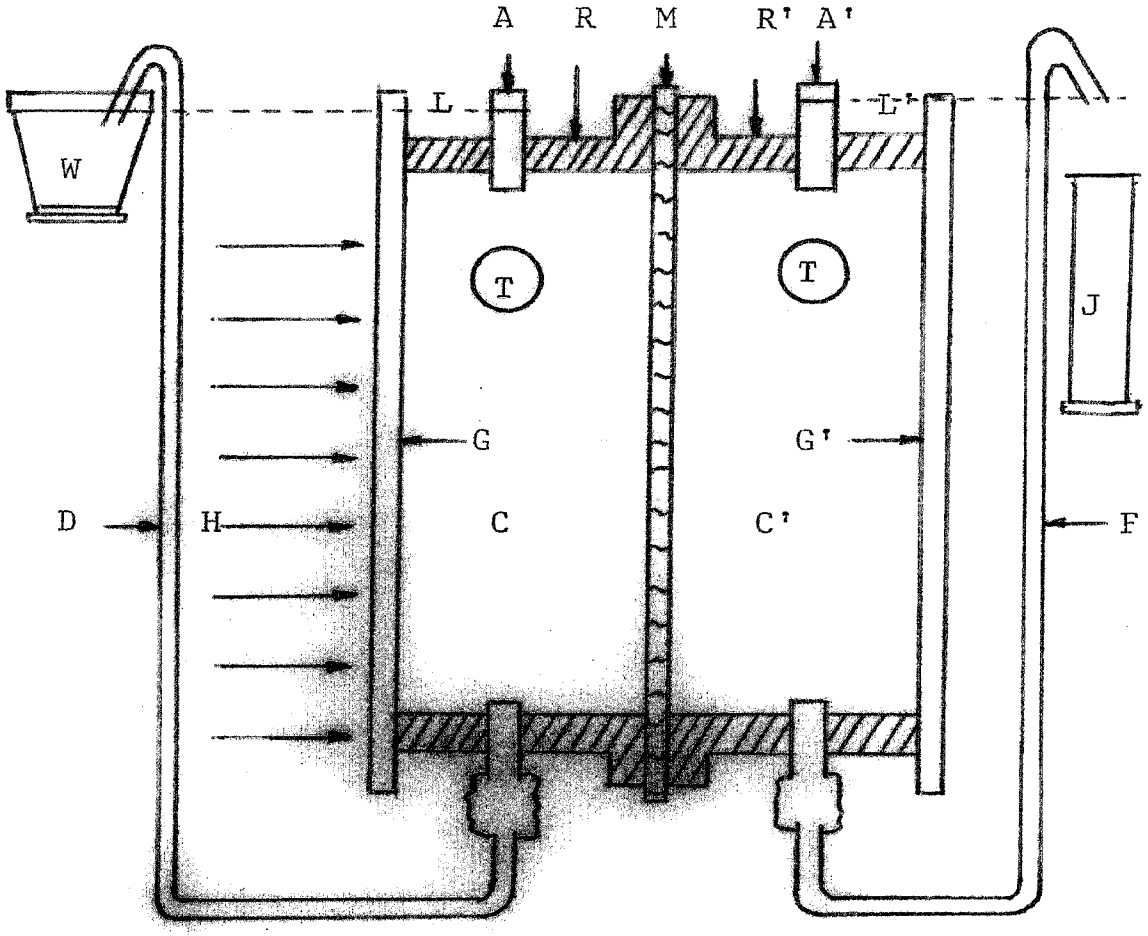
C. Study of Membrane Life:

As discussed before, membrane life controls not only the membrane cost, but also the operating labor of replacement. Therefore, membrane life plays a major part in economic considerations. A study of membrane life is discussed below.

The apparatus used for determining the membrane life is shown in Figure 3.5. A continuous supply of heat to the salt water maintains the salt water temperature sufficiently high to ensure the transfer of water vapor from the hot salt water through the pores of membrane to the cold fresh water. The depletion of the water component in the salt water is continuously replenished by fresh water from the water reservior, so that the salt concentration is assumed to be constant. In order to ensure that the salt concentration remained constant, salt water is replaced once a week. The product water is received by a graduated cylinder.

The desalination system is operated with a very small liquid flow rate and in open to the atmosphere. In order to reduce the evaporation of both salt water and fresh water to the atmosphere to a negligible extent, the temperature used for this study is below $140^{o}F$.

The membranes used are made from a dispersion of 1.0 gram of Owens-Corning Fiberglas A-A, 0.5 ml of DuPont Teflon Dispersion 30 B for the first, and 0.2 ml for the second, 4 ml of 5% aluminum sulfate for 113 sq. cm of membrane. The aluminum sulfate is used to improve teflon retention in the fiber mat. Five drops of



A, A' Vents and Filling Tubes
H Heat from Light Bulb
C, C' Chambers for Salt and Fresh Water
D Make-up Line
F Fresh Water Outlet
G, G' Glass Plates
J Graduated Cylinder
L, L' Liquid Levels
R Rubber Gaskets
T Thermometers
W Water Reservoir

Fig. 3.5 Experimental Apparatus

glacial acetic acid are added to aid dispersion. After formation by filtration, the membrane is pressed, dried and baked at about $600^{o}F$ to provide reasonable strength and suitable water repellency for desalination.

The mass transfer of water through the membrane is correlated with the pressure driving force across the membrane based on equation 2.11.

Rewriting equation 2.12,

$$N_A = K_m (aP_A - P_{A2}) \tag{3.64}$$

where N_A is the rate of mass transfer of water, lb./hr. sq. ft., a is the activity of water in salt water, P_A is the water vapor pressure at the bulk salt water temperature, in. Hg., P_{A2} is the water vapor pressure at the bulk water temperature, in. Hg., K_m is the overall mass transfer coefficient, lb./hr. sq. ft. in. Hg.

P_A and P_{A2} can be evaluated from the equation given in Appendix B, and activity a remains approximately constant at 0.96 in the range investigated.

Let F be the total mass transfer in lb. through the effective membrane area, A_m in sq. ft., in a time interval t hr. Then,

$$K_m = \frac{N_A}{(aP_A - P_{A2})}$$

$$= \frac{F}{A_m \, t \, (aP_A - P_{A2})} \tag{3.65}$$

K_m values calculated from equation 3.65 were plotted against time as shown in Figure 3.6. Either Side 1 or Side 2 represents the membrane side which is in contact with hot salt water during an experimental run.

When the membrane becomes less effective for desalination after a certain time, the membrane is flushed in place and the contents of each side and the heat flows are reversed, so that mass transfer is in the opposite direction.

The decrease in membrane performance is believed to be due to salt or other solids on the surfaces and in the pores of the membrane, which act as obstacles for mass transfer. The membrane is regenerated by reversing heat flow and mass transfer. In this way, salt solids can be cleaned out by water vapor condensing after being transferred through the pores. However, new salt or other solids will gradually cover the other surface. Finally, the membrane can not be regenerated satisfactorily after an extensive operating history. The results on the first membrane are shown in Figure 3.6. K_m values after 140 days are always below 0.25 lb./hr. ft.^{2} in. H_g and appear unsatisfactory, due to a rapid decrease in performance. A membrane life of four months with an average K_m value of 0.3 lb./hr. ft.^{2} in. Hg appears to be reasonable based on Figure 3.6.

However, it is probably necessary to reverse the membrane operation frequently. If membrane performance is allowed to decrease to very low values, the pores or surfaces are nearly

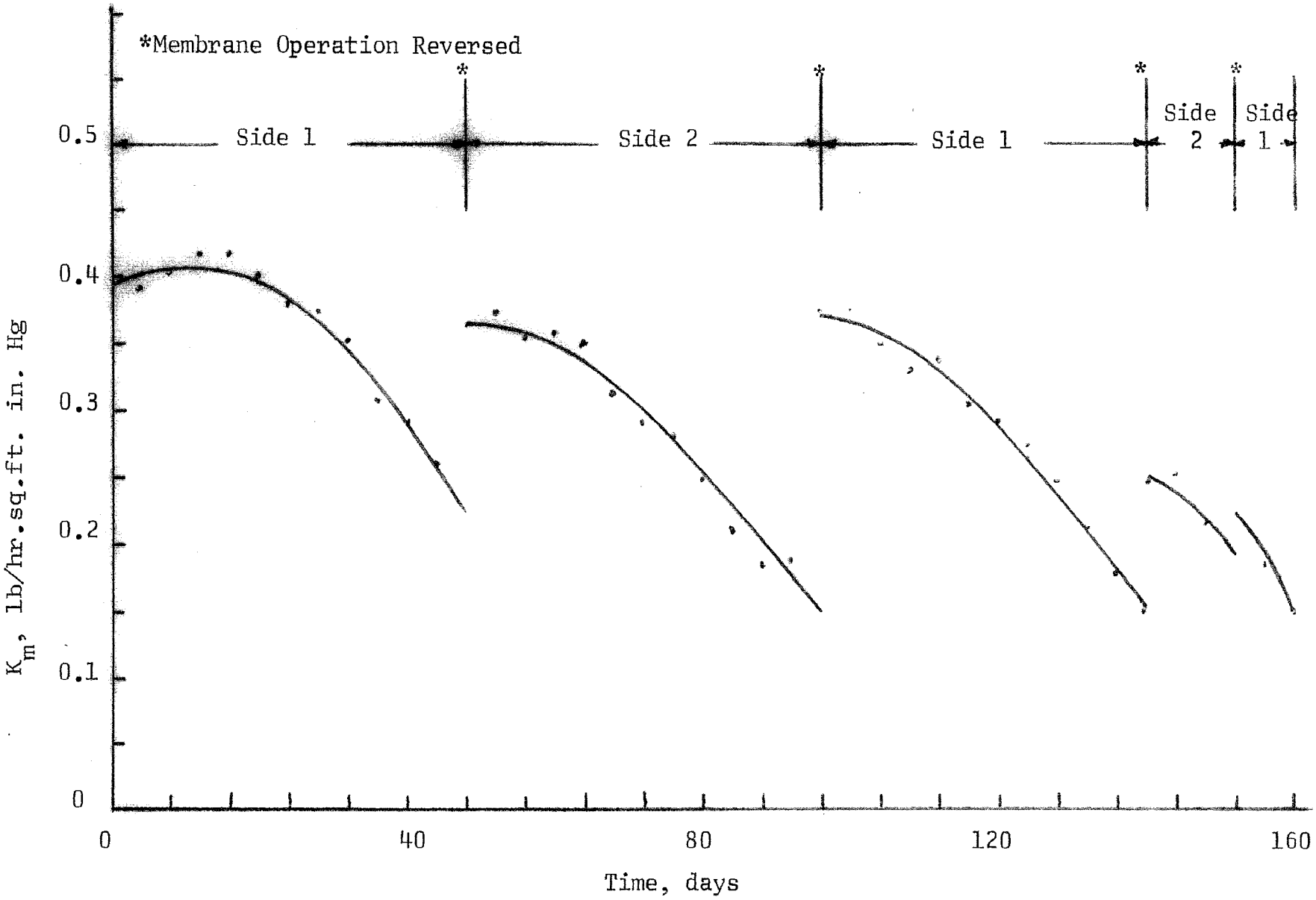


Fig. 3.6 Overall Mass Transfer Coefficient, K_m, vs. Time

FIRST MEMBRANE

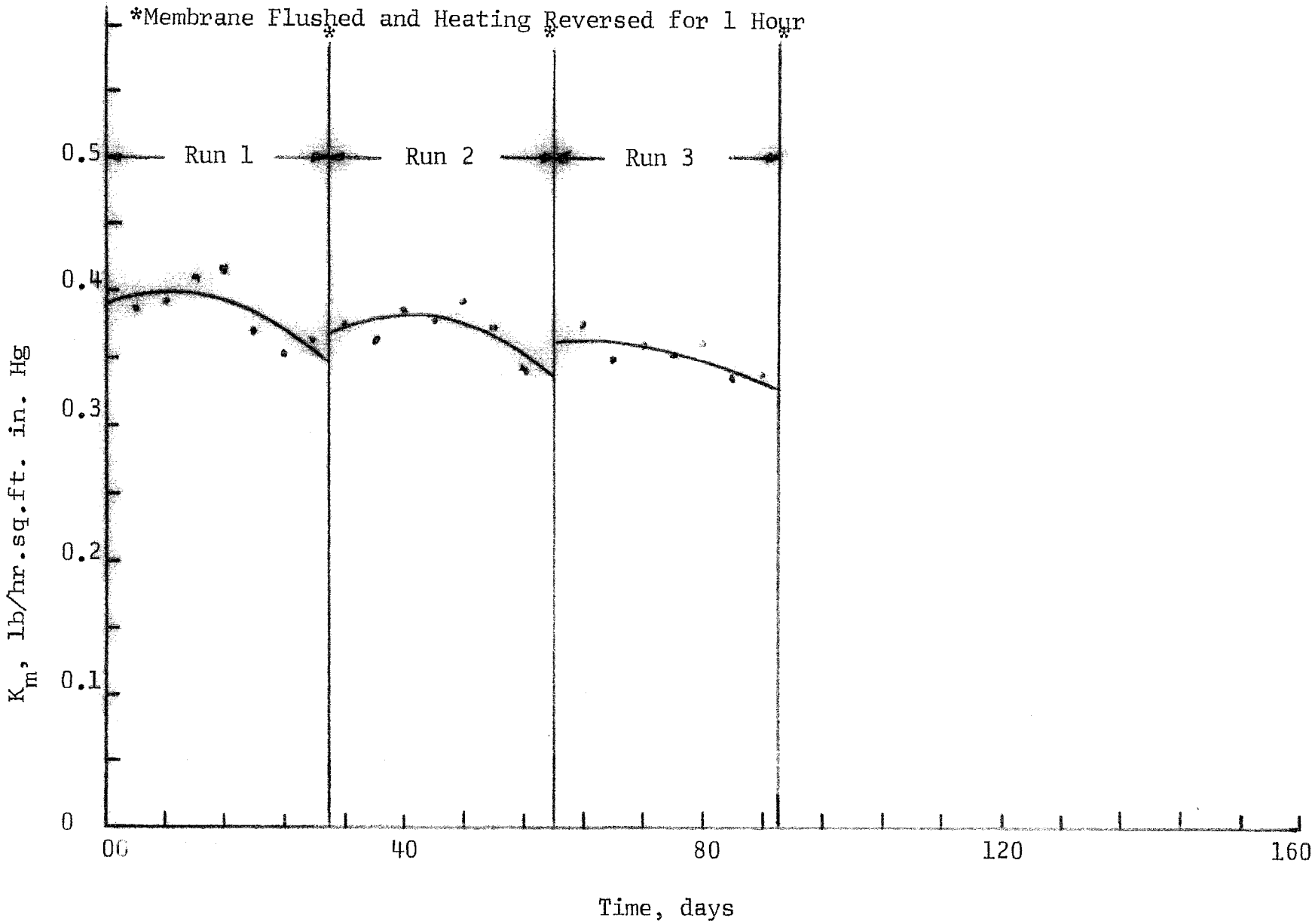


Fig. 3.7 Overall Mass Transfer Coefficient, K_m, vs. Time

SECOND MEMBRANE

completely filled with salt or other solids, and the membrane is difficult to regenerate.

A second membrane was used with fixed salt water and fresh water sides. The membrane performance, after a slight decrease, was regenerated by filling both sides with fresh water and then reversing heat flow for one hour. The water vapors are thus transferred in the opposite direction, and clean the pores, but they deposit no new solids from the fresh water.

The operation time for each run was 30 days. The results plotted in Figure 3.7 show that this method can effectively regenerate the membrane performance, in a similar manner as used with the first membrane, even though there is apparently some difference in the two membranes. The ability to regenerate membranes is believed to be increased as the duration of the reversed heat flow is increased and as the operating cycle time decreases.

In either method, the surface and pores of the membrane can be cleaned to regenerate the membrane without the necessity of removing the membrane, but in the second method it is not necessary to switch the contents on each side of the membrane. In the first method, the membrane is operating as it is regenerated.

It was not the purpose of this investigation to evaluate regeneration methods, but tentatively the second method appears to be preferable, both from the point of view of membrane performance and the purity of product water.

In order to be conservative, a membrane life of four months at an average K_m of 0.3 lb./hr. ft.^{2} in. Hg obtained from Figure 3.6 is used for the study of the economics of these water desalination techniques.

IV. RESULTS AND DISCUSSION

A. Method I:

1. The Effect of Temperature Driving Force on Water Conversion and Required Area:

Membrane area required and water conversion rate can be calculated from equation 3.6 and equation 3.7. Based on a saline water flow of W_{s1} = 1000 lb./hr. entering the evaporation at T_{s1} = 500°F, the water conversion and required membrane area were plotted in Figure 4.1 and Figure 4.2 as functions of outlet salt water temperature at various temperature driving forces D_T, where $D_T = T_s - T_c$. The water conversion rate W_c, can be obtained as $W_c = 1000 - W_{s2}$. Where W_{s2} is the saline flow leaving the evaporator at a specified temperature T_{s2}.

As shown in Figure 4.1, the temperature driving force D_T has only little effect on the conversion as a function of outlet salt water temperature. For D_T values between 5 and 30°F the curves lie evenly spaced between the upper and the lower line. The reason for this lack of effect of D_T is that the conversion is mainly a function of the sensible heat lost by the sea water.

The requirement of membrane area for desalination as shown in Figure 4.2 is sharply affected by the temperature driving force, especially at low temperatures. The lower the temperature driving force, the smaller the pressure driving force generated for desalination, and the more the membrane area is required.

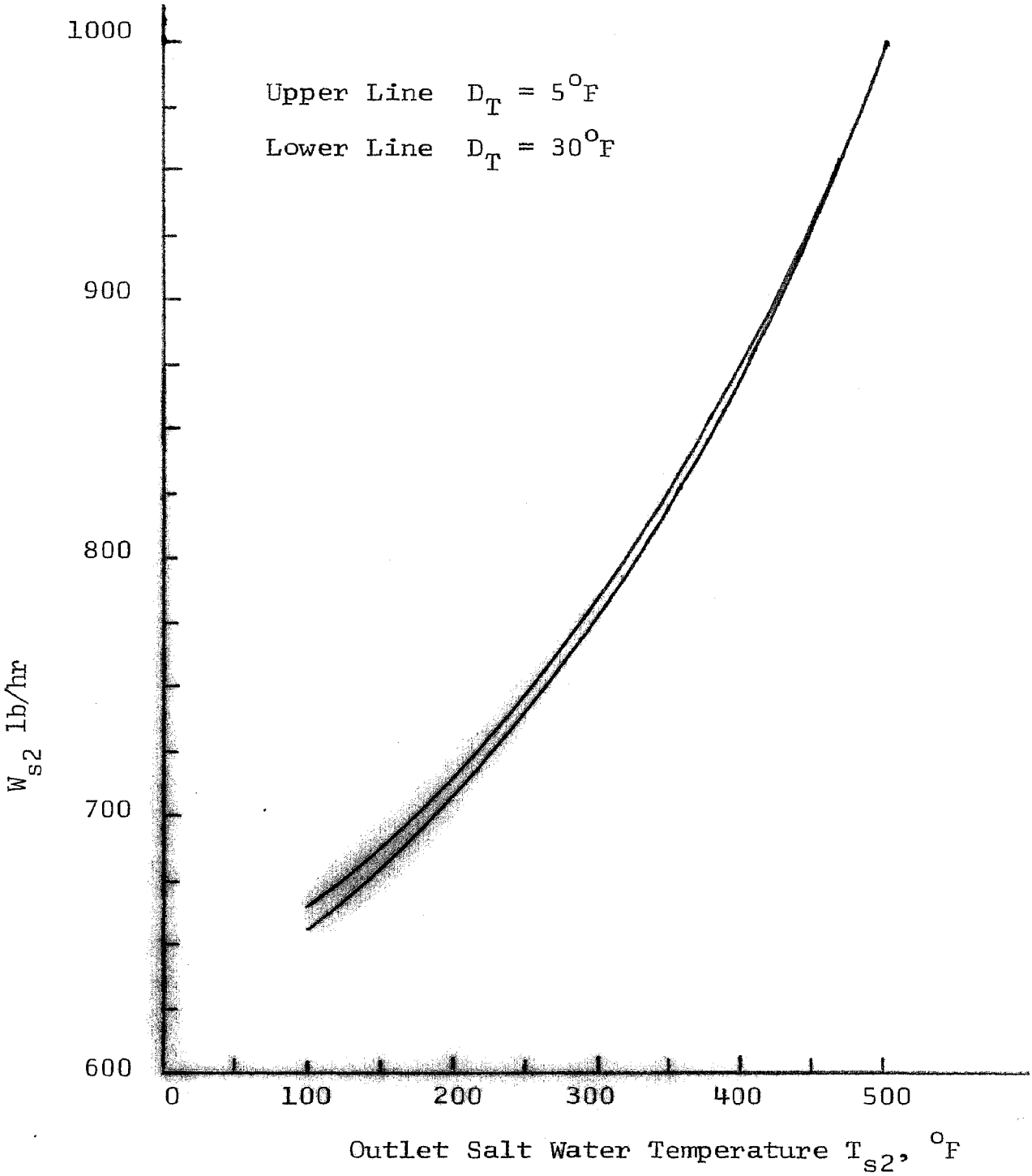


Fig. 4.1 Sea Water Flow W_{s2} vs. Sea Water Temperature, T_{s2}, leaving the Evaporator-Condenser

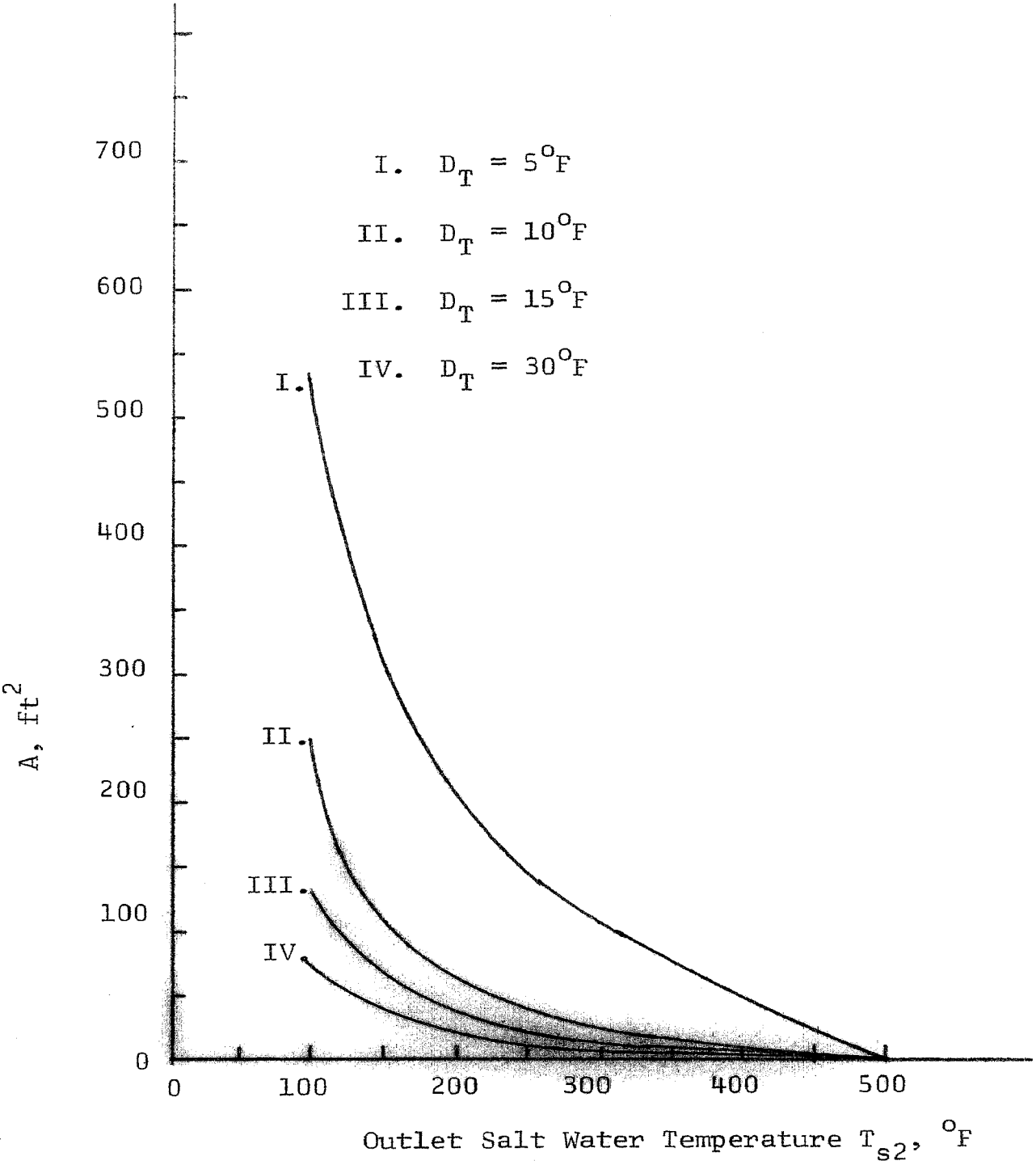


Fig. 4.2 Membrane Area Required A vs. Sea Water Temperature, T_{s2}, leaving the Evaporator-Condenser

2. Water Cost:

Equation 3.18 through equation 3.28 with the aid of equation 3.6 and equation 3.7 are used to evaluate all the necessary items of equipment and operating cost. Water cost is then calculated from equation 3.29 and equation 3.30. A plot of the results was used to visually select an optimum cost and design.

Figure 4.3, 4.4, 4.5 and 4.6 are plots of water cost vs. sea water temperature leaving the membrane evaporator-consdenser at a constant temperature driving force, D_T in the evaporator with the temperature difference, D_{TEX} in the heat exchanger as a parameter.

At constant D_T, larger D_{TEX} will decrease the cost of heat exchanger, but increase the cost of heat. At constant D_{TEX}, larger D_T will decrease the cost of evaporator and membrane, but increase the cost of heat. The effects of D_T and D_{TEX} on the water cost can be inspected from these water cost figures. The optimum cost is approximately 60 cents per 1000 gallons with the optimum operation conditions approximately $D_T = 15^oF$, $D_{TEX} = 20^oF$, and $T_{s2} = 200^oF$ as shown in Figure 4.5. This means that sea water feeds in the evaporator at 500^oF and leaves at 200^oF.

Because the optimum outlet temperature is about 200^oF, these results indicate that below 200^oF it is better not to utilize the heat in the exit sea water for further evaporation, but to utilize such heat to bring up the entering stream temperature by a heat exchanger.

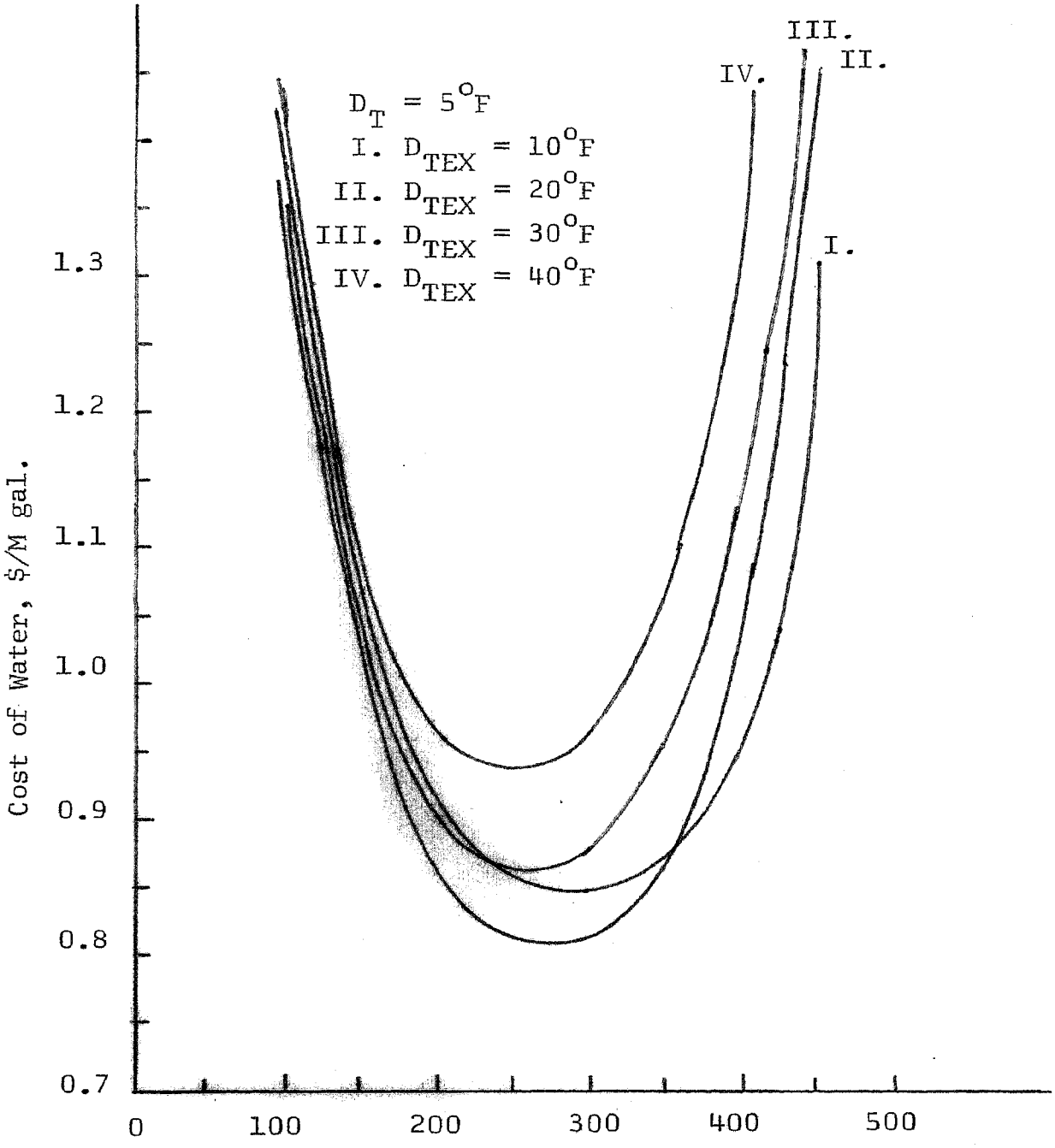


Fig. 4.3 Cost of Water Per M gal, vs. Sea Water Temperature Leaving the Membrane Evaporator-Condenser

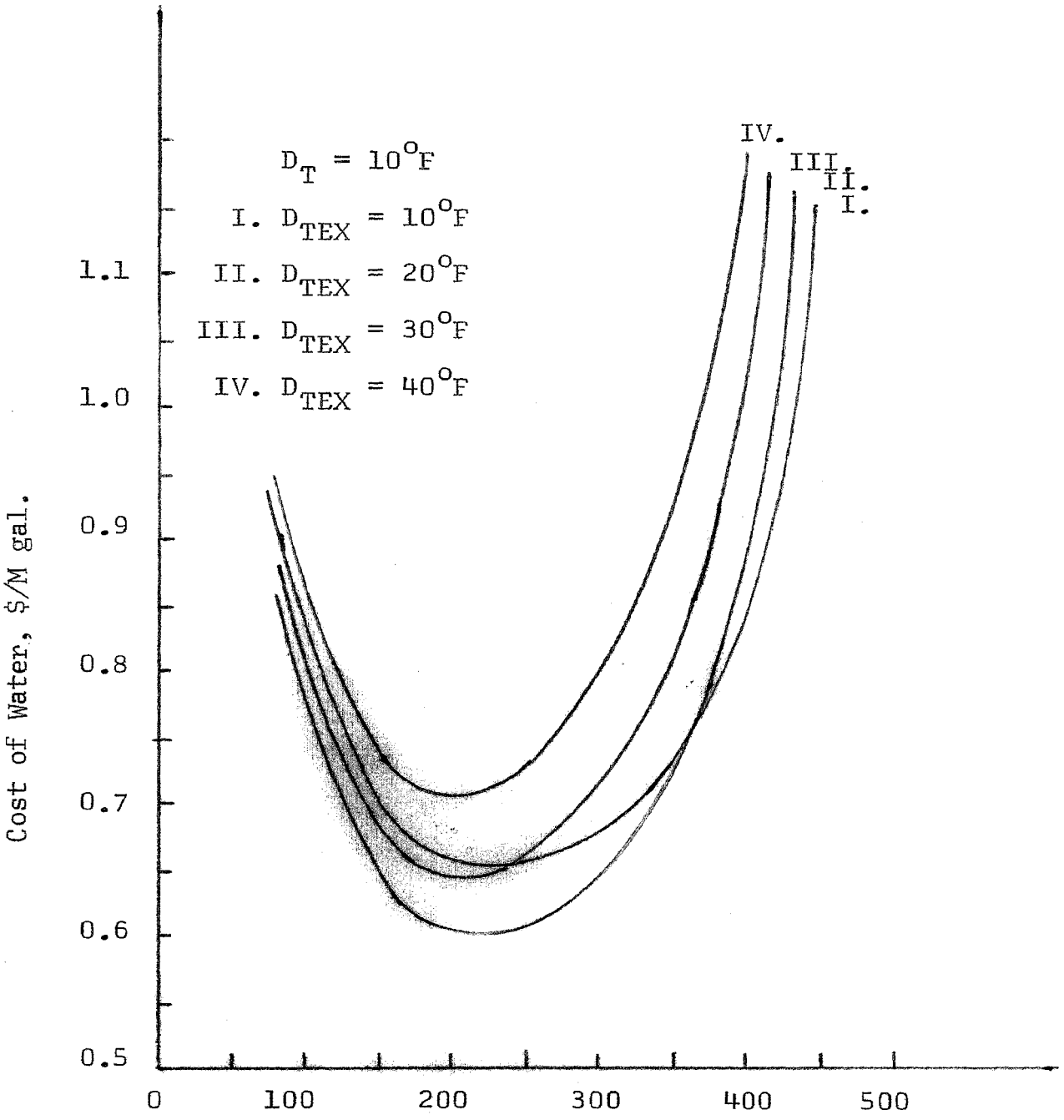


Fig. 4.4 Cost of Water Per M gal, vs. Sea Water Temperature Leaving the Membrane Evaporator-Condenser

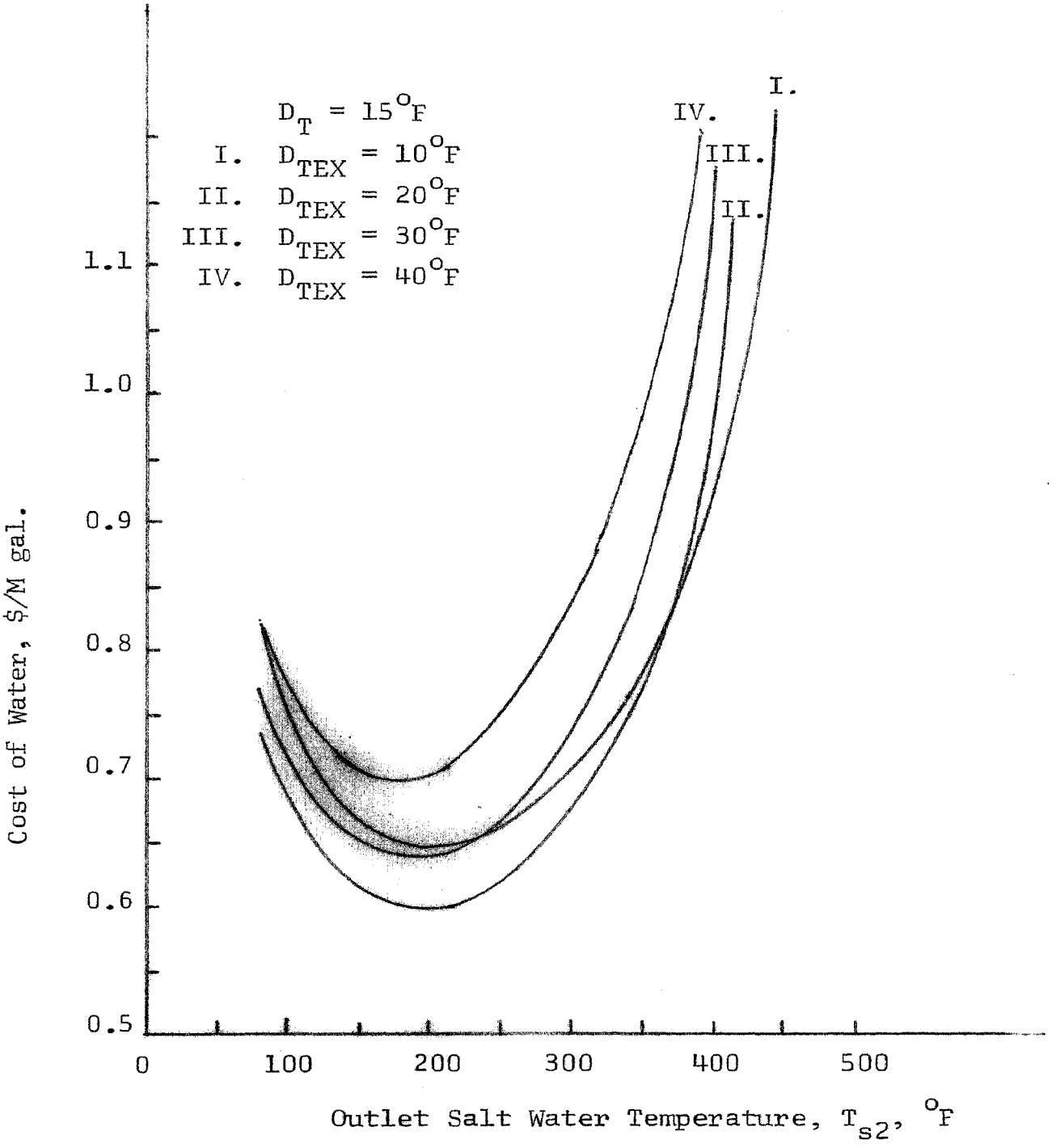


Fig. 4.5 Cost of Water Per M gal. vs. Sea Water Temperature Leaving the Membrane Evaporator-Condenser

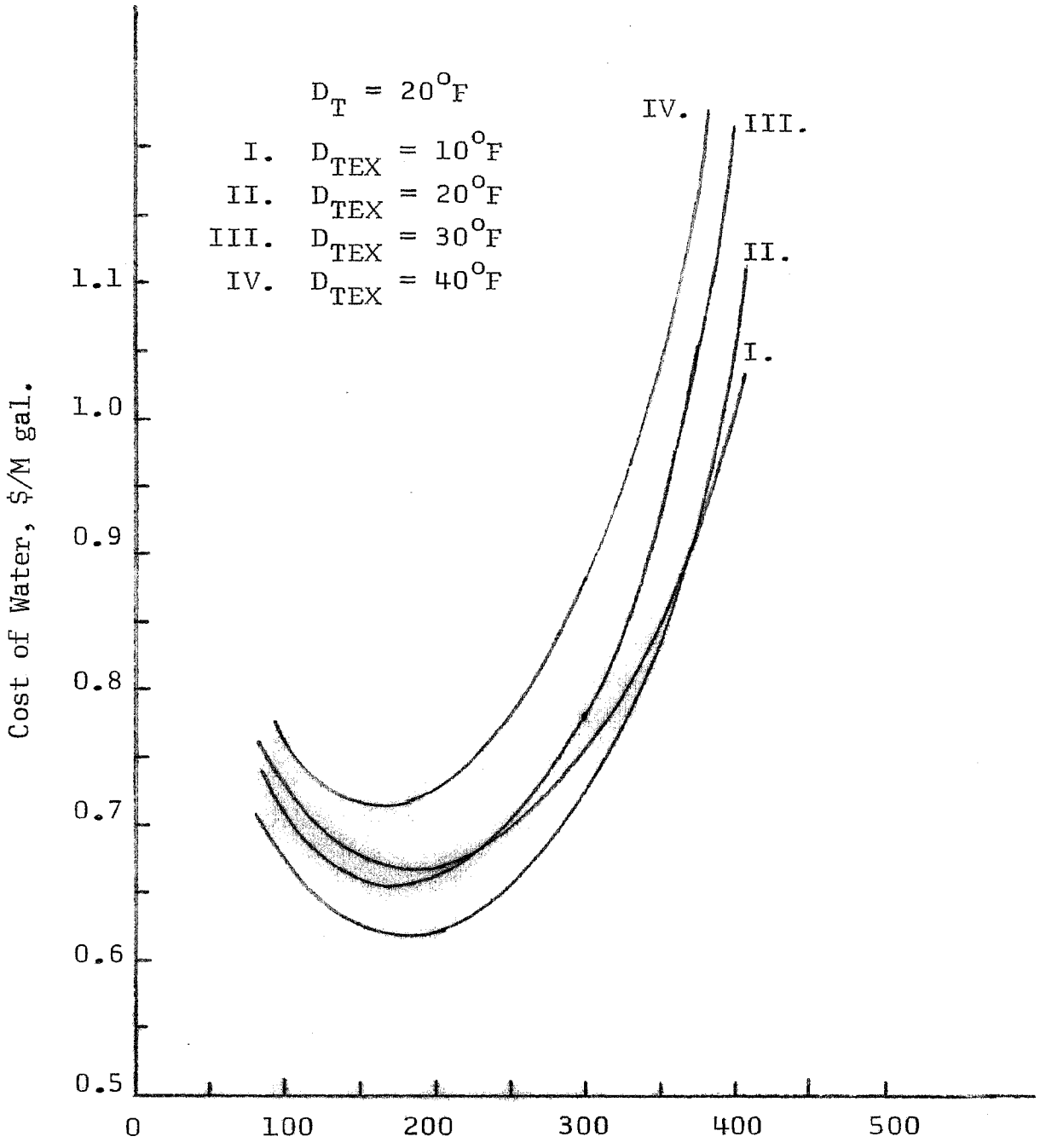


Fig. 4.6 Cost of Water Per M gal. vs. Sea Water Temperature Leaving the Membrane Evaporator-Condenser

If the salt water stream leaving the evaporator is reduced to about 100°F by further evaporation, the preheater is not required, but considerably more membrane area is required, as shown in Figure 4.2. The evaporation from 200°F down to 100°F could be carried out at atmospheric pressure, and in a different type of apparatus the cost per unit area of membrane could probably be reduced. However, this aspect was not studied in further detail because the primary purpose of this investigation was to determine the most economical region of operation and optimum temperature differences.

A detailed analysis of plant costs and operating costs is given in Appendix E on the results with sea water from 500°F to 100°F in the evaporator and with D_T = 15°F and D_{TEX} = 20°F and an overall cost of 69.3 cents per 1000 gallons of water product.

A plot of approximate optimum cost vs. K_m is given in Appendix H to estimate optimum cost if a better membrane performance is obtained in the near future.

3. <u>The Effect of Saline Water Temperature Entering the Evaporator on Water Cost</u>:

As discussed before, with constant temperature driving force across the membrane, the pressure driving force for desalination will be greater at higher temperatures. Therefore, the optimum saline temperature entering the evaporator should be its maximum possible temperature.

TABLE 4.1 COST COMPARISON BETWEEN VARIOUS SALINE TEMPERATURES ENTERING THE EVAPORATOR, METHOD I

	Saline temperature, T_{s1}, entering the evaporator		
	$500^{o}F$	$350^{o}F$	$210^{o}F$
T_{s2}, ^{o}F	100	100	100
D_T, ^{o}F	15	15	15
D_{TEX}, ^{o}F	20	20	20
C_{OEV}, MM$	2.606	4.891	11.455
C_{OE1}, MM$	2.018	3.343	8.367
C_{OE3}, MM$	0.104	0.104	0.104
C_{OHR}, MM$	0.424	0.702	1.756
C_{OPM}, MM$	0.150	0.026	-
C_{OEG}, MM$	0.120	0.024	-
C_{OMO}, MM$	0. 039	0.010	-

Continued

C_{MB}, $/day	628	1,179	2,761
C_{HT}, $/day	2,338	3,867	9,679
C_{PO}, $/day	387	71.6	-
Capital Cost,			
$ per gal. per day	1.06	1.74	4.11
Water Cost,			
$ per M gal.	0.693	1.087	2.56

A comparison of water costs between various saline entering temperatures is shown in Table 4.1 based on the conditions, T_{s2} = 100°F, D_T = 15°F and D_{TEX} = 20°F. The notation of the cost items is defined in the Nomenclature.

At high saline water entering temperature, the pump, engine, motor and power cost more due to the larger pressure requiring more energy to transport the sea water to the operation pressure. All the other costs increase, as expected, as the saline entering temperature decreases. The costs of pump, engine, motor and power consumption at 210°F entering temperature are assumed to be of negligible importance as compared with the other costs.

B. Method II:

1. Water Conversion and the Required Length of Evaporator:

The equations and method for the calculations of water conversion rate and the required length of evaporator have been discussed in the Design Equation for Membrane Evaporator-Condenser of Method II. The relationship between sea water flow, W_s, and its leaving temperature, T_{s2}, is plotted in Figure 4.7, based on the following conditions at the hot end of the evaporator-condenser:

(1) 1000 lb./hr. of sea water entering 2 in. membrane tube in 4 in. evaporator at 500°F, and

(2) fresh sea water feed leaving eight 3/4 in. heat exchanger tubes at 470°F.

Figure 4.7 can also be applied to estimate this relationship

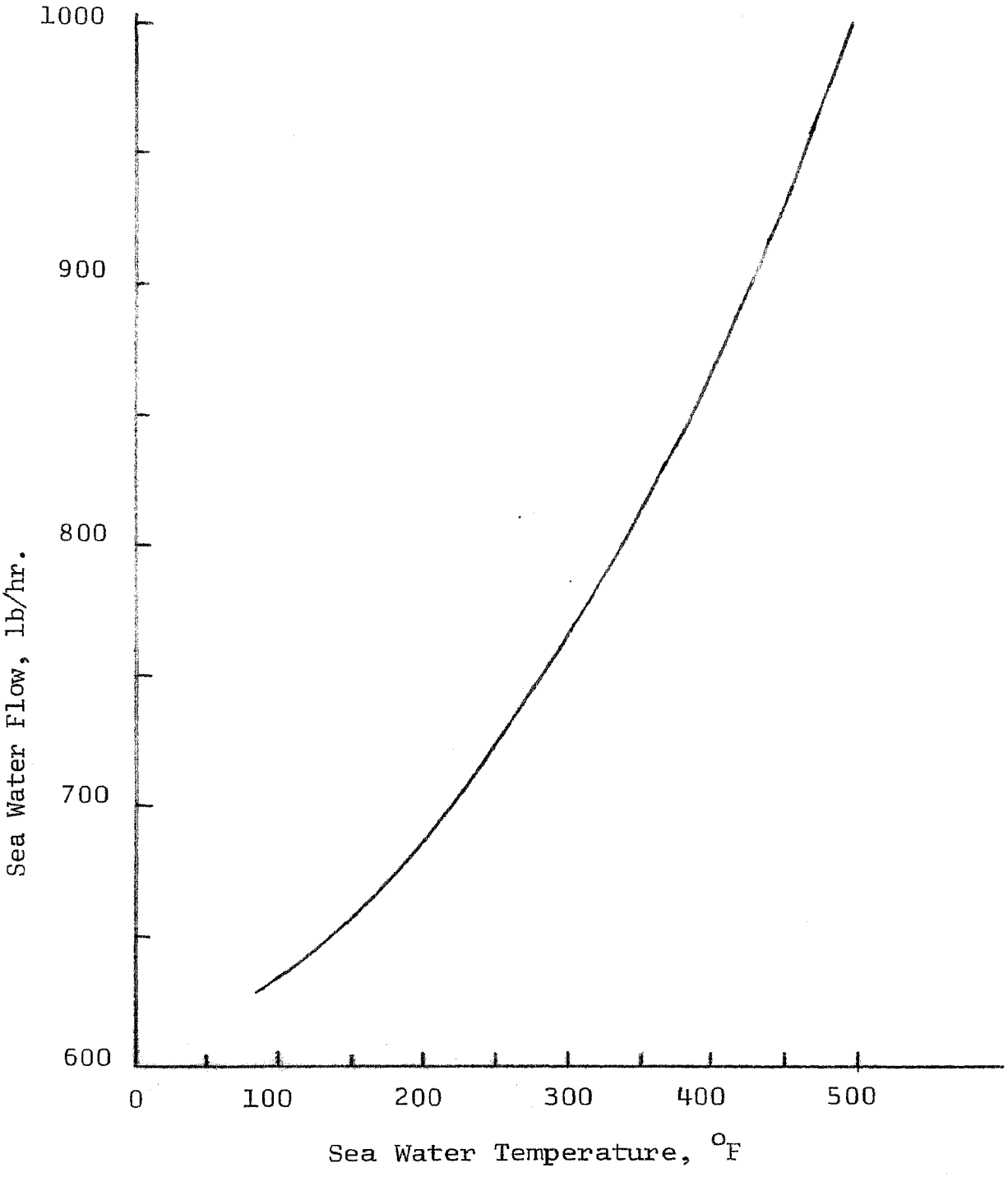


Fig. 4.7 Sea Water Flow in the Membrane Tube vs. Sea Water Temperature with P_M = 0.523 ft., P_E = 1.570 ft. and T_{s31} = 470°F.

within 1.0% error at the following conditions:

(a) $0.523 \leqslant P_m \leqslant 1.088$ ft.

(b) $0.393 \leqslant P_E \leqslant 2.093$ ft.

(c) $440 \leqslant T_{s31} \leqslant 490^{o}F$.

Again, the conversion is primarily dependent on the sensible heat lost by the evaporating stream. Within the above practical limits, various shapes of membrane tubes and various combinations of heat exchanger tubes on the shell side of a 4 in. evaporator are given in Table 4.2. Each shape other than circular is only given with the heat exchanger tubes which have a maximum perimeter for heat transfer.

A combination of the variables in equation 3.42 has only little effect on sea water flow and water conversion rate at a certain temperature of sea water leaving the evaporator.

Curve I, II and III in Figure 4.8 through 4.15 represent the salt water temperature in the membrane tube, product water temperature on the shell side, and fresh salt water temperature in the heat exchanger tubes, respectively. The perimeter P_m of membrane tube, the perimeter P_E of heat exchanger tubes and the fresh salt water temperature T_{s31} leaving the heat exchanger tubes have much influence on the required length of evaporator. The effect of heat exchanger tube perimeter P_E is shown in Figure 4.8, 4.9 and 4.11, and that of membrane tube perimeter P_m is shown in Figure 4.11, 4.13 and 4.14. The optimum P_E and P_m are their practical maximum

TABLE 4.2 VARIOUS SHAPES OF MEMBRANE TUBE IN COMBINATION WITH HEAT EXCHANGER TUBES ON THE SHELL SIDE OF 4 INCH EVAPORATOR

Membrane Tubes			Heat Exchanger Tubes		
Diameter, in.	Shape	Perimeter P_M, ft.	Diameter, in.	Number	Perimeter P_E, ft.
2	Circular	0.523	3/4	4	0.785
2	Circular	0.523	3/4	6	1.177
2	Circular	0.523	3/4	8	1.570
2	Circular	0.523	1/2	16	2.093
4	Equilateral Triangular	0.866	1/2	16	2.093
4	4 Pointed Star	1.088	1/2	16	2.093
4	6 Pointed Star	1.154	1/2	12	1.570

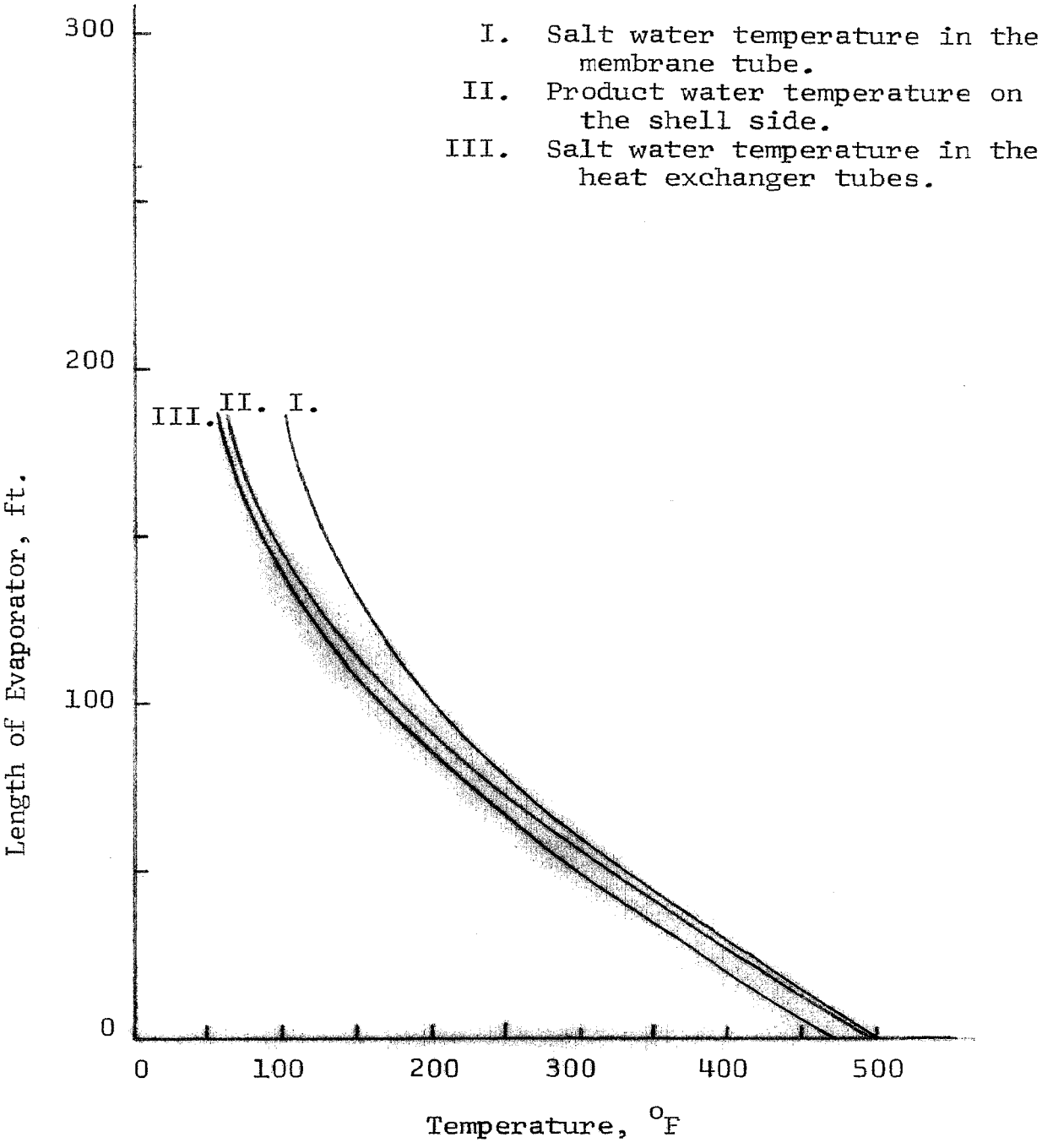


Fig. 4.8 Length of Evaporator vs. Temperature with P_M = 0.523 ft., P_E = 0.785 ft., and T_{s3L} = 470°F

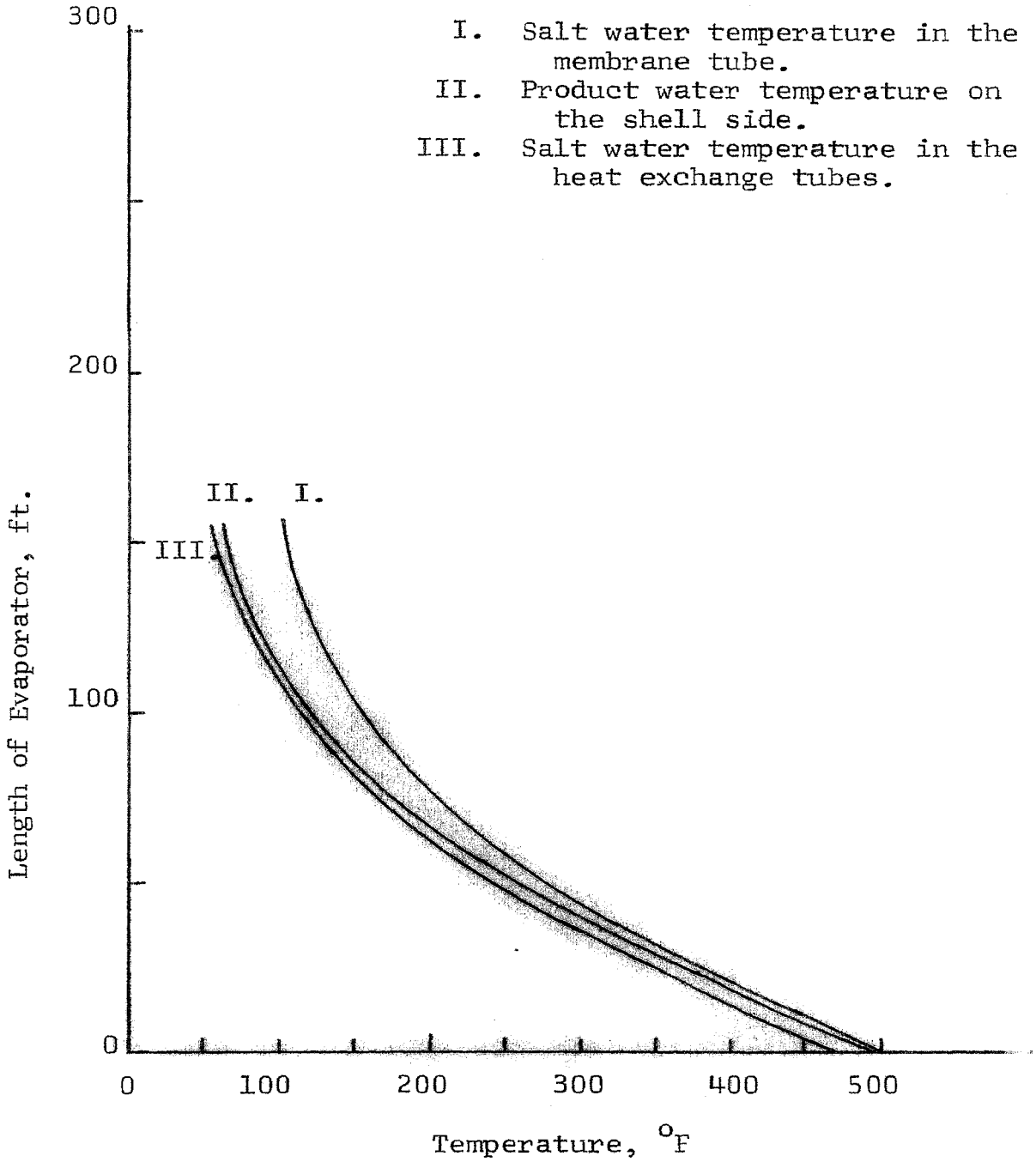


Fig. 4.9 Length of Evaporator vs. Temperature with P_M = 0.523 ft., P_E = 1.177 ft., and T_{s31} = 470°F

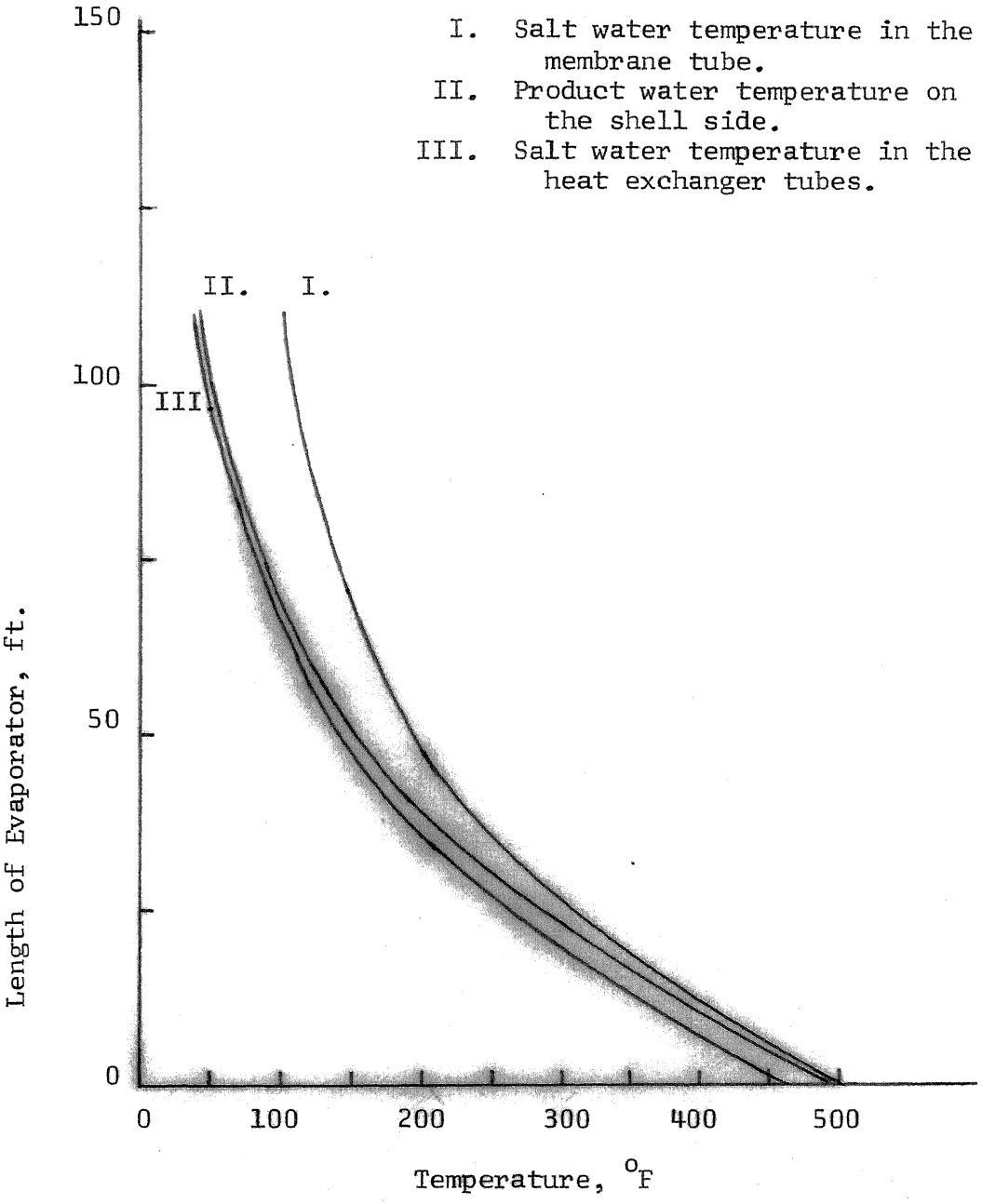


Fig. 4.10 Length of Evaporator vs. Temperature with P_M = 0.523 ft., P_E = 1.570 ft., and T_{s31} = 460°F

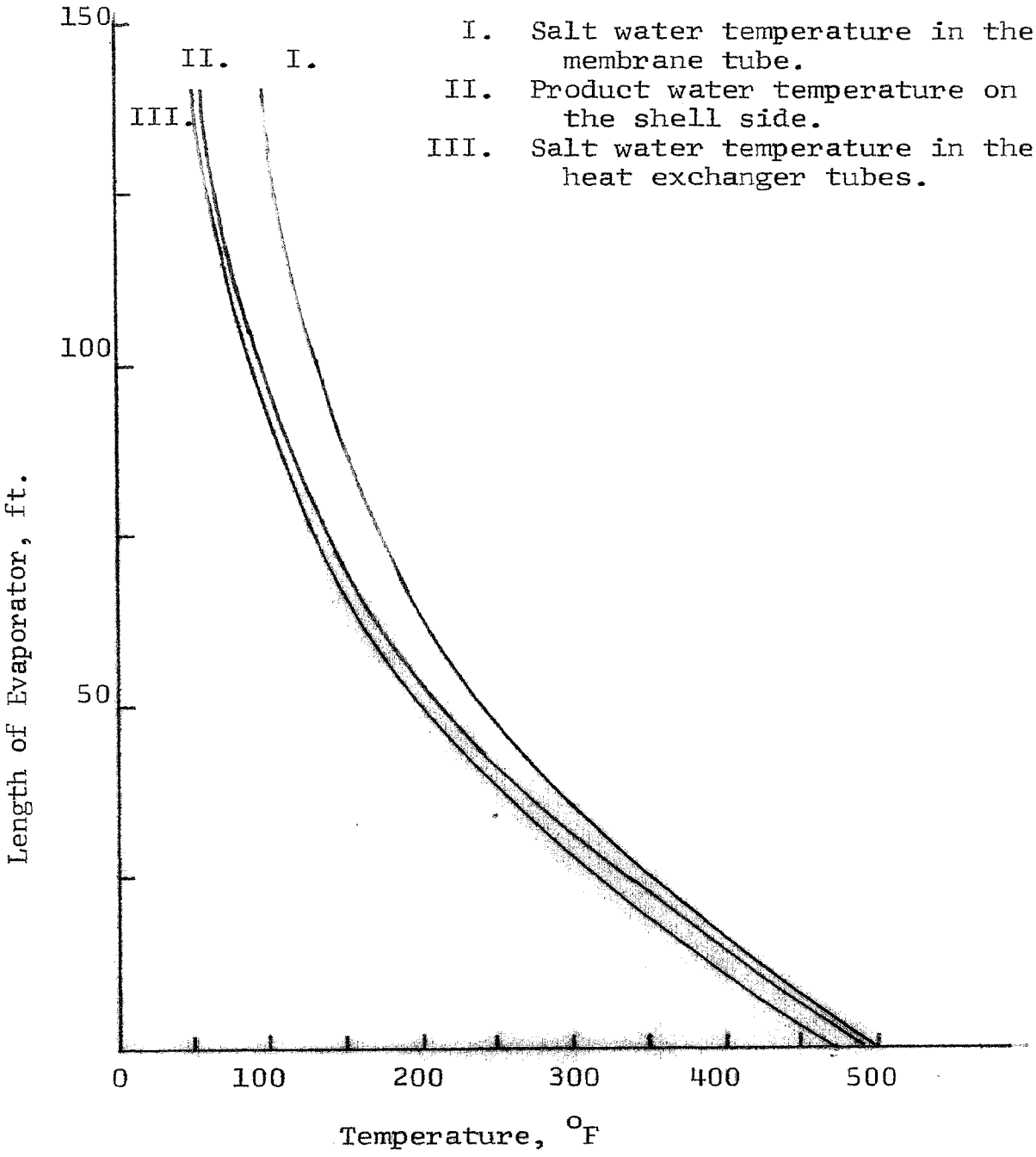


Fig. 4.11 Length of Evaporator vs. Temperature with P_M = 0.523 ft., P_E = 1.510 ft., and T_{s31} = 470°F

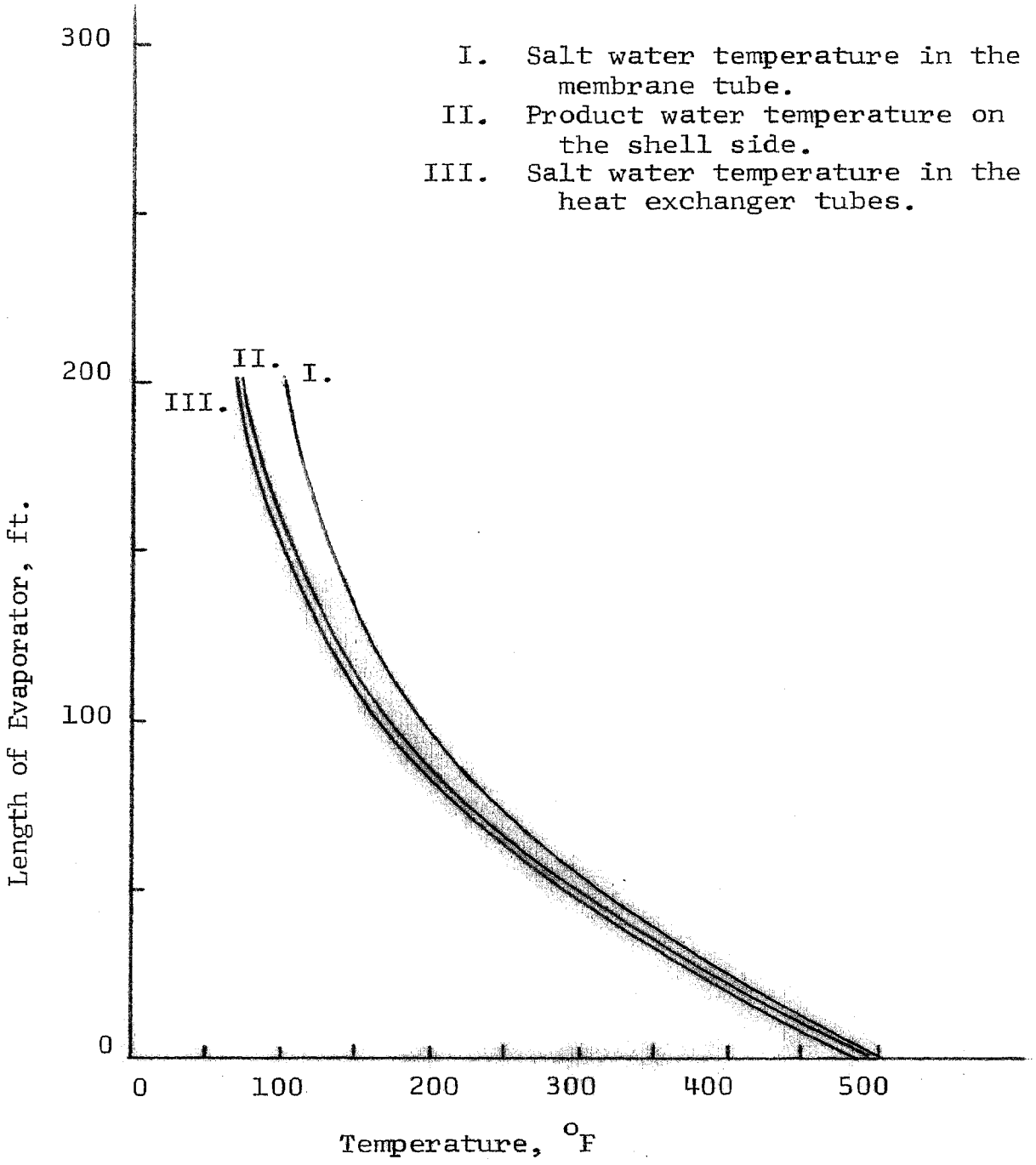


Fig. 4.12 Length of Evaporator vs. Temperature with P_M = 0.523 ft., P_E = 1.576 ft., and T_{s31} = 480°F

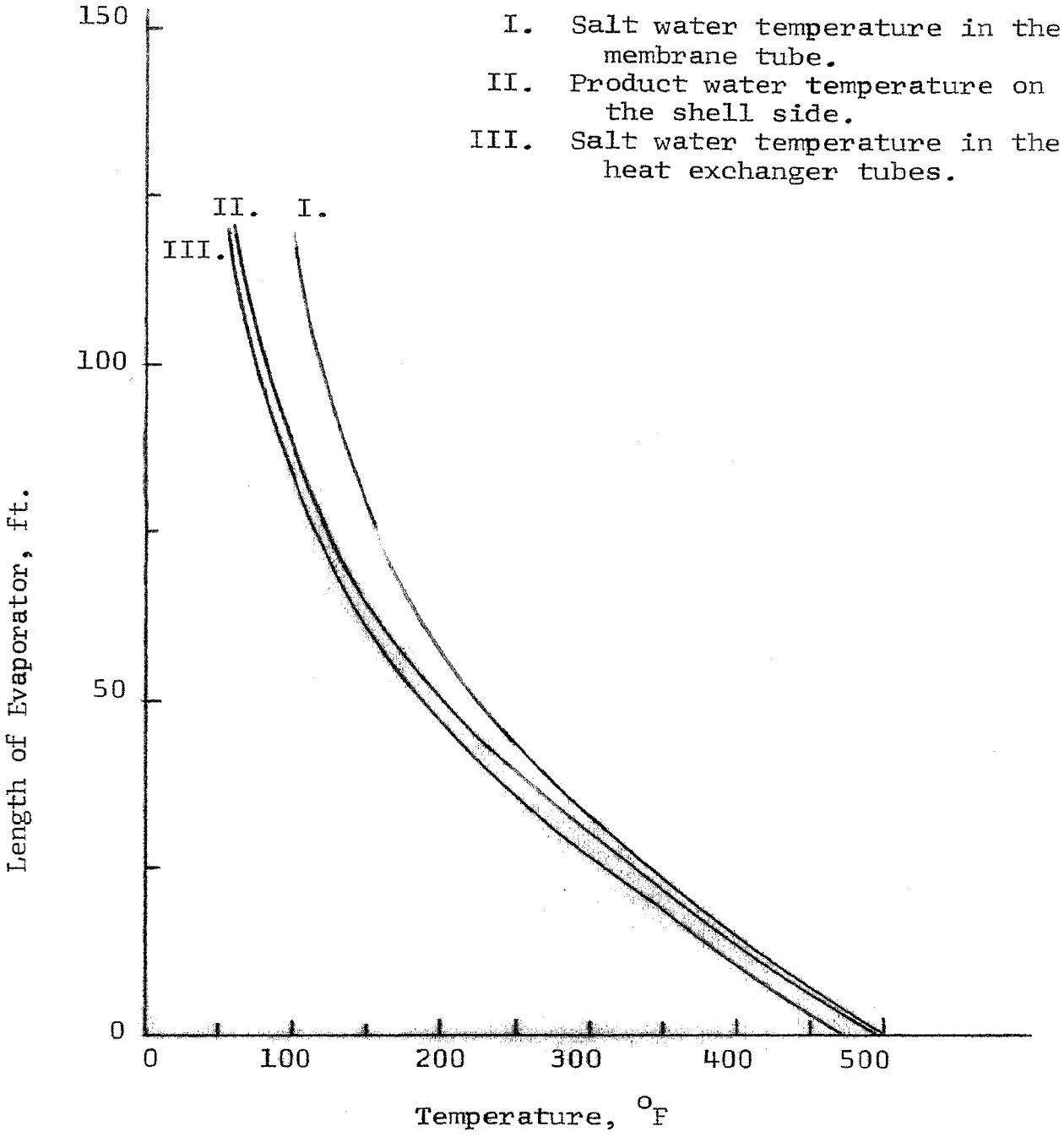


Fig. 4.13 Length of Evaporator vs. Temperature with P_M = 0.667 ft., P_E = 1.570 ft., and T_{s31} = 470°F

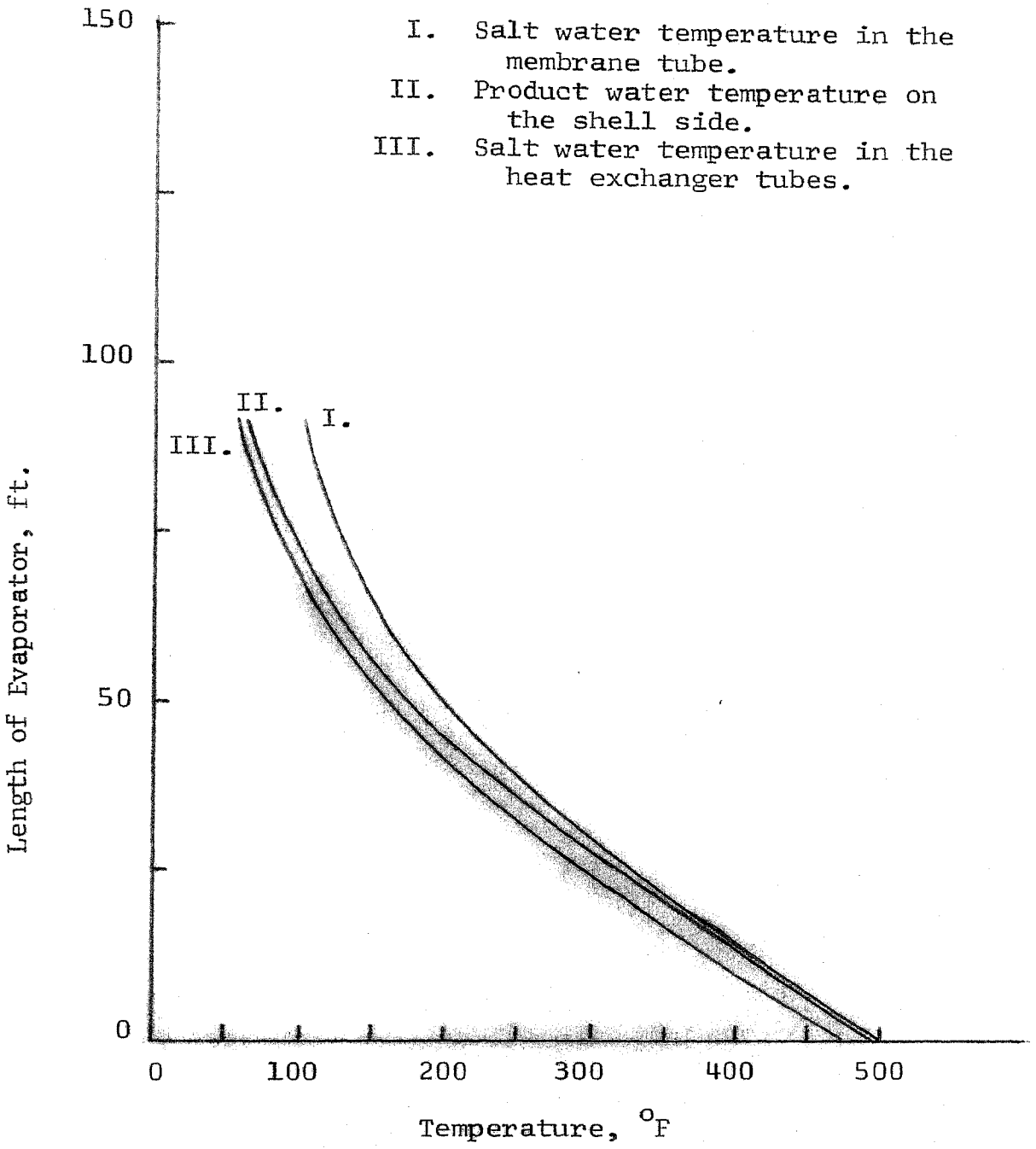


Fig. 4.14 Length of Evaporator vs. Temperature with P_M = 1.088 ft., P_E = 1.570 ft., and T_{s31} = 470°F

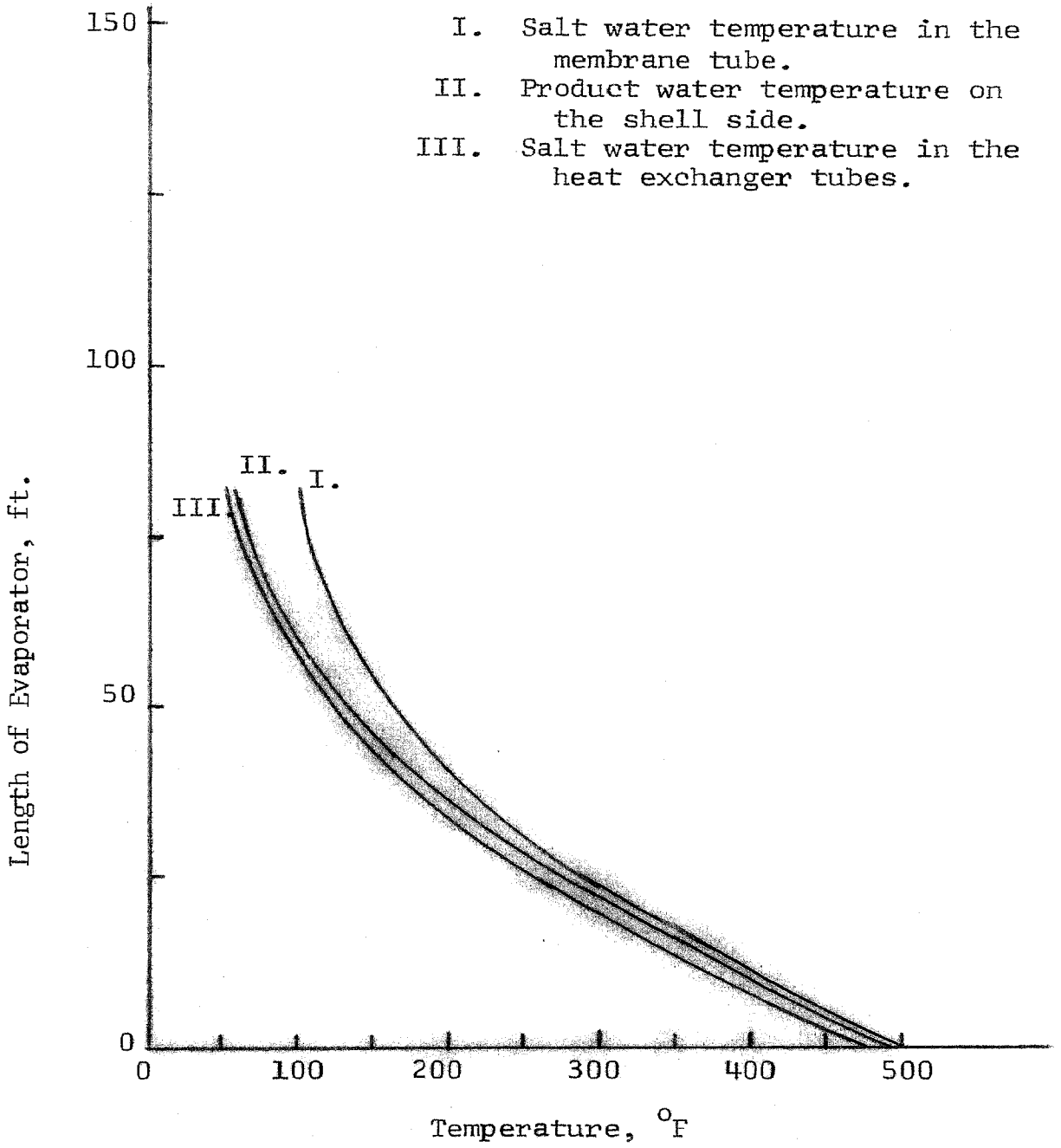


Fig. 4.15 Length of Evaporator vs. Temperature with P_M = 1.088 ft., P_E = 2.093 ft. and T_{s3I} = 470°F

possible in a 4 in. evaporator. A membrane tube of 4 pointed star shape (P_m = 1.088 ft.) with sixteen 1/2 in. heat exchanger tubes (P_E = 2.093 ft.) on the shell side of a 4 in. evaporator is considered to be optimum. Lower salt water temperatures leaving the heat exchanger tubes increase the temperature driving force across the membrane and therefore require less evaporator length as shown in Figure 4.10, 4.11 and 4.12.

2. Water Cost:

The equations and method for the evaluation of water cost have been discussed before. The results are plotted in Figure 4.16 through 4.21. Curve I, II and III in these water cost figures represent the fresh sea water temperature, T_{s31}, leaving the heat exchanger tubes at 480, 470, and 460^{o}F, respectively. The effect of the membrane tube perimeter P_m, the total heat exchanger tube perimeter P_E, and the fresh sea water leaving temperature, T_{s31}, is indicated on Figure 4.16 to 4.21 by means of changing one variable and keeping the other two constant. The optimum value of P_m and P_E are their maximum values if other variables are constant. The maximum P_m provides sufficient area for mass transfer and decreases the requirement of evaporator length. The maximum P_E provides enough area for heat transferred from product water to salt water being heated in the heat exchanger tubes. The temperature of the product water is, therefore, maintained as low as possible which introduces the maximum temperature driving force across the membrane for mass transfer. The approximate optimum water cost as shown in Figure 4.21 is 47.2 cents per

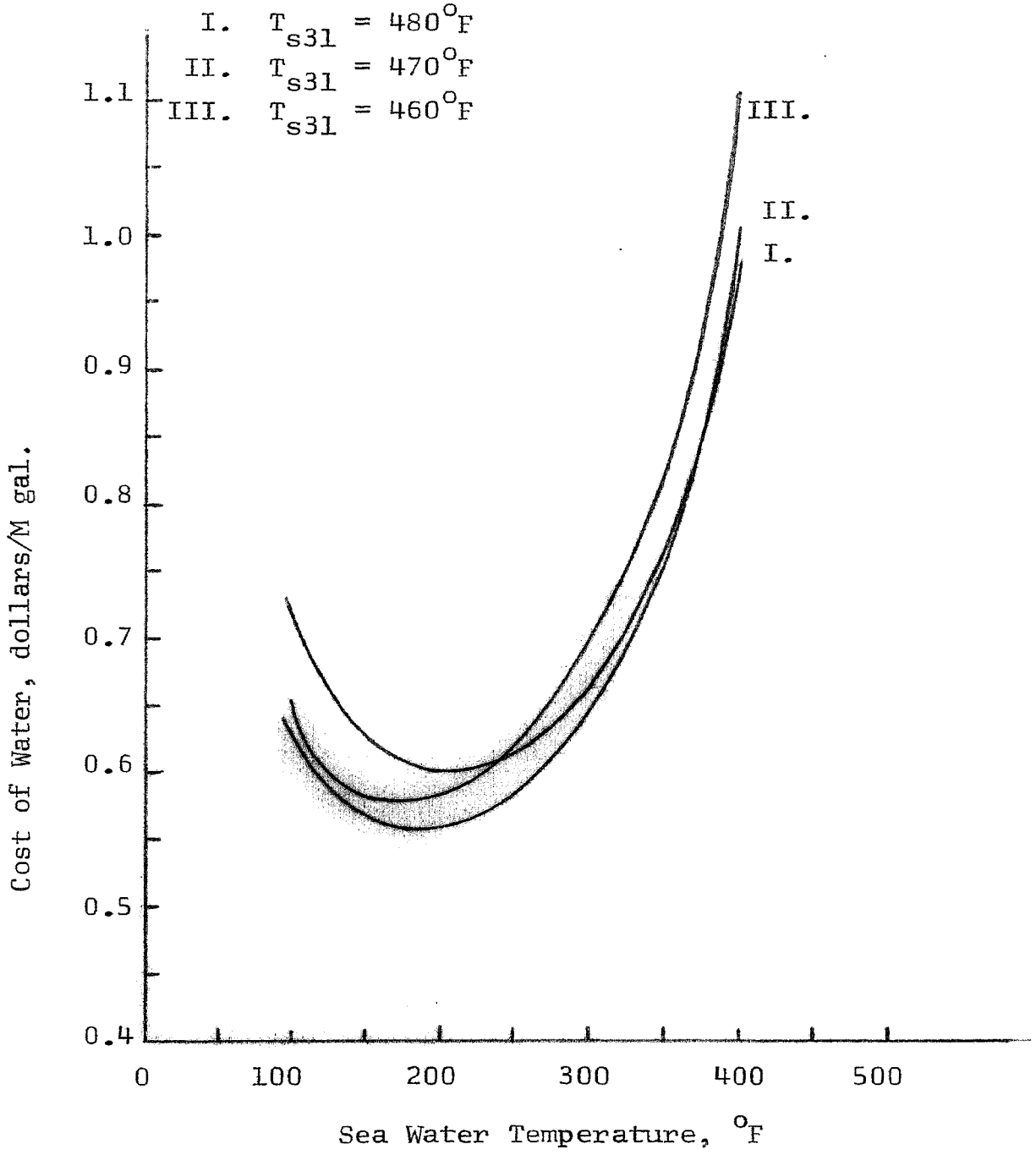


Fig. 4.16 Cost of Water Per M Gal. vs. Sea Water Temperature with P_M = 0.523 ft. and P_E = 0.785 ft.

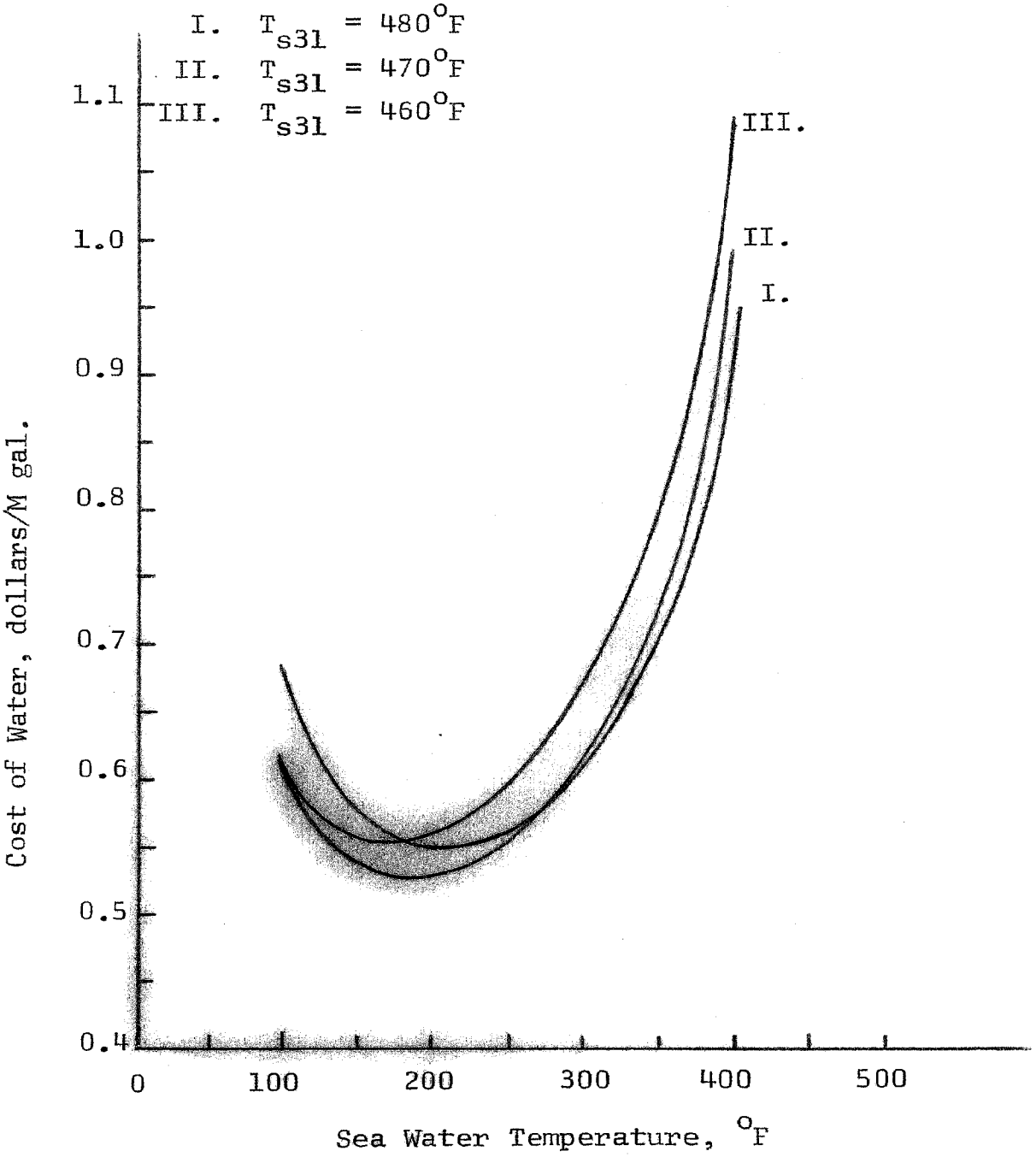


Fig. 4.17 Cost of Water Per M Gal. vs. Sea Water Temperature with P_M = 0.523 ft. and P_E = 1.177 ft.

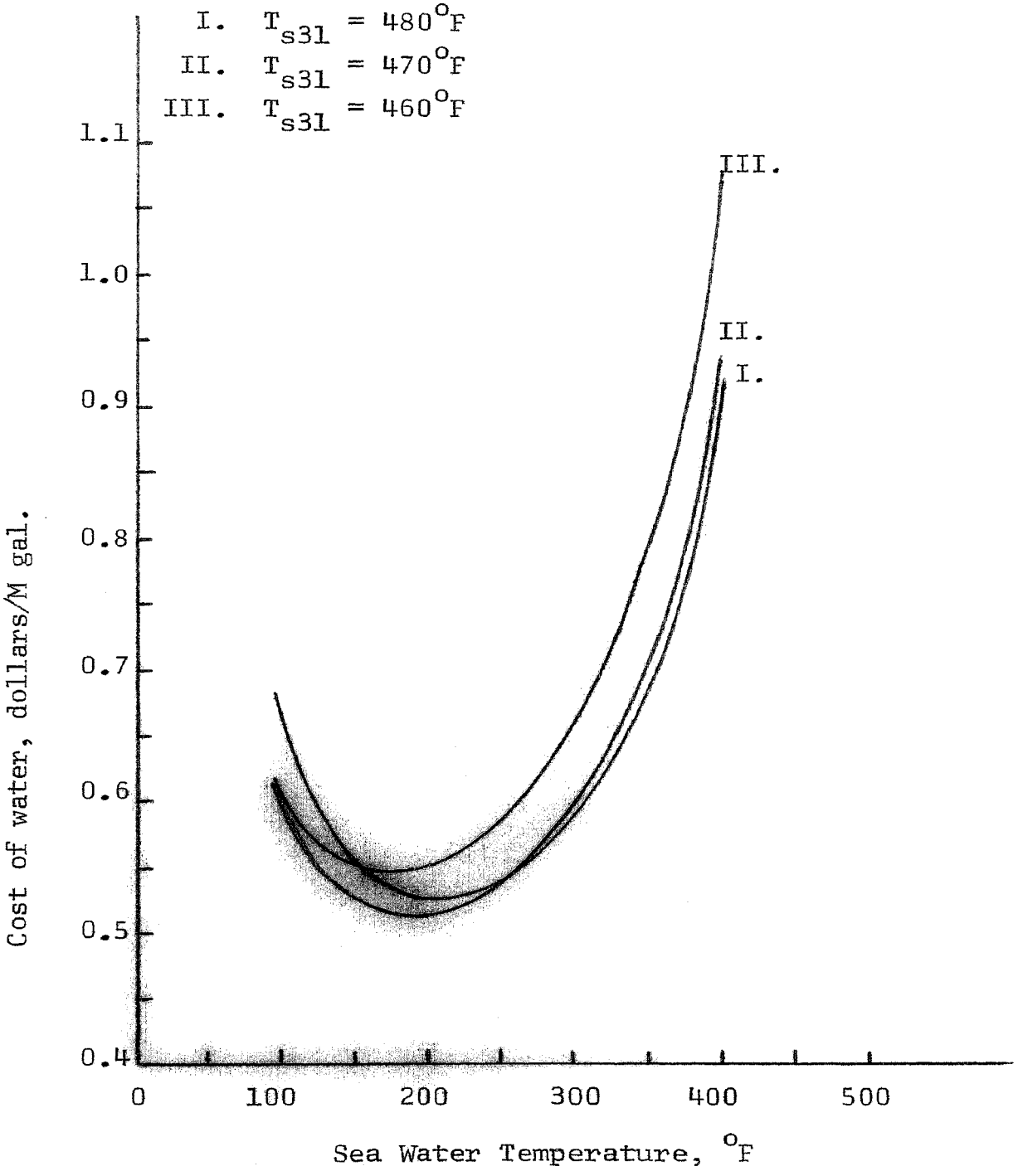


Fig. 4.18 Cost of Water Per M Gal. vs. Sea Water Temperature with P_M = 0.523 ft. and P_E = 1.177 ft.

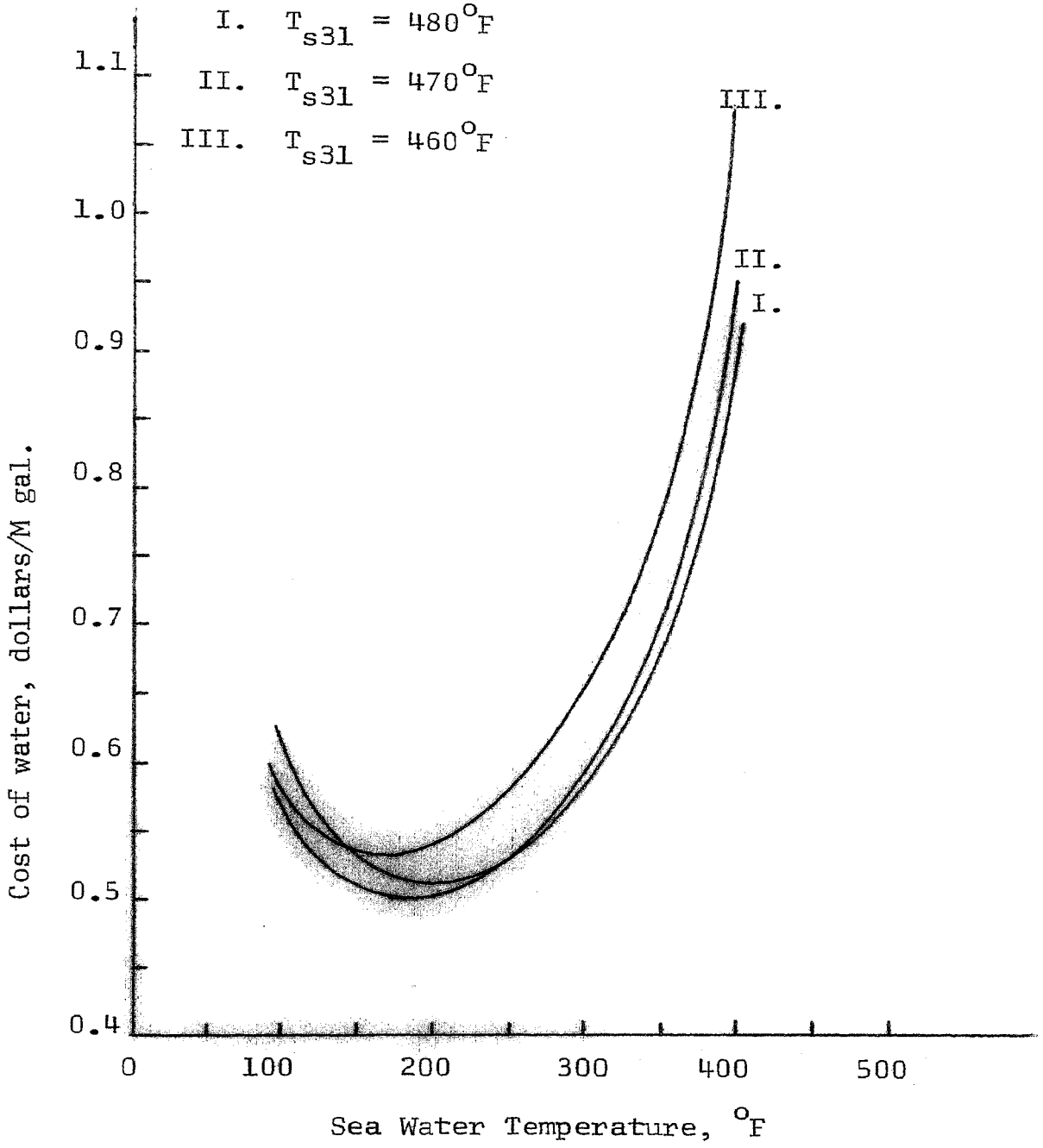


Fig. 4.19 Cost of Water Per M Gal. vs. Sea Water Temperature with P_M = 0.667 ft. and P_E = 1.570 ft.

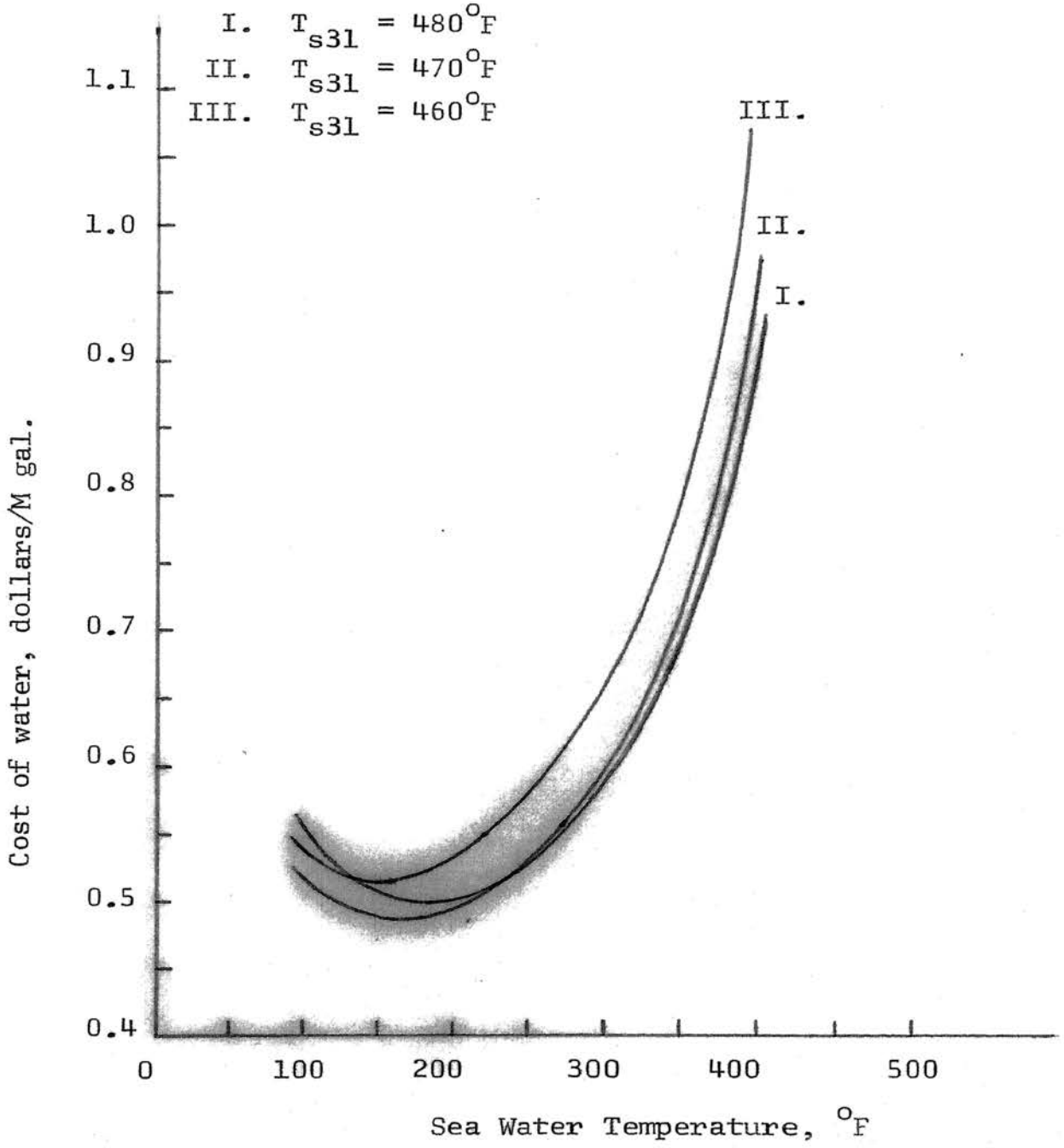


Fig. 4.20 Cost of Water Per M Gal. vs. Sea Water Temperature with P_M = 1.088 ft. and P_E = 1.570 ft.

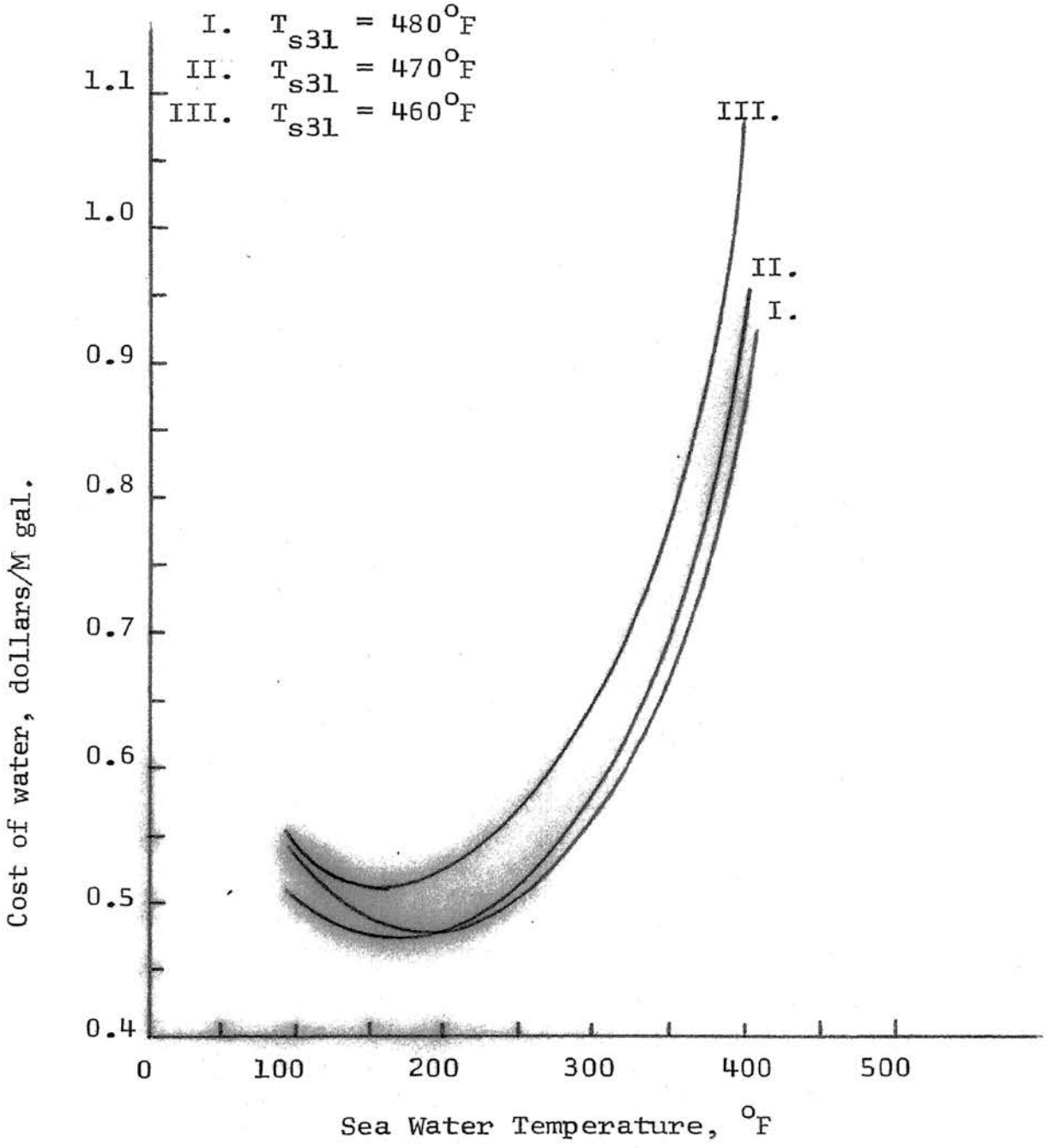


Fig. 4.21 Cost of Water Per M Gal. vs. Sea Water Temperature with P_M = 1.088 ft. and P_E = 2.093 ft.

1000 gallons water product with T_{s2} = 150°F, T_{s31} = 470°F, P_m = 1.088 ft. and P_E = 2.093 ft. In a 4 in. evaporator, a membrane tube of four pointed star shape with sixteen 1/2 in. heat exchanger tubes on the shell side was considered to meet these perimeter requirements. A survey of T_{s31} at temperatures near 470°F (using increments of 2°F) also shows that T_{s31} at 470°F is the actual optimum point.

A detailed analysis of plant costs and operating costs is given in Appendix F on the results with T_{s1} = 500°F, T_{s2} = 130°F, T_{s31} = 470°F, P_m = 1.088 ft. and P_E = 2.093 ft., and an overall cost of 47.7 cents per 1000 gallons of water product.

A plot of approximate optimum cost vs. K_m is given in Appendix H to estimate optimum cost if a better membrane performance is obtained in the near future.

3. <u>The Effect of Saline Temperature Entering the Evaporator on Water Cost</u>:

Based on the conditions, T_{s2} = 130°F, P_m = 1.088 ft. and P_E = 2.093 ft., a cost comparison between saline water temperatures, T_{s1}, entering the evaporator is shown in Table 4.3. Cost of pump, engine, motor and power consumption at various T_{s1} were assumed equal to that of Method I. The same conclusion was obtained in this method as in Method I; that is the optimum saline temperature entering the evaporator is at its maximum possible value.

TABLE 4.3 COST COMPARISON BETWEEN VARIOUS SALINE TEMPERATURES ENTERING THE EVAPORATOR, METHOD II

	Saline temperature, T_{s1}, entering the evaporator		
	$500^{o}F$	$350^{o}F$	$212^{o}F$
T_{s31}, ^{o}F	470	320	182
T_{s2}, ^{o}F	130	130	130
C_{OEV}, MM$	2.370	3.755	8.071
C_{OEX}, MM$	0.042	0.113	0.411
C_{OHR}, MM$	0.352	0.604	1.756
C_{OPM}, MM$	0.150	0.026	-
C_{OEG}, MM$	0.120	0.024	-
C_{OMO}, MM$	0.039	0.010	-
C_{MB}, $/day	324	614	1104
C_{HT}, $/day	1942	3174	9674
C_{PO}, $/day	387	71.6	-

Continued

Capital Cost,			
$ per gal. per day	0.615	1.002	1.988
Water Cost,			
$ per M gal.	0.477	0.727	1.720

C. Discussion and Comparison of Method I and II:

The countercurrent flow system described in Method I utilizes effectively the possible advantage that the latent heat of water vapor evaporated from the hot salt water side can be transferred to the product water similar to a countercurrent heat exchanger. The outlet temperature of product water can be theoretically elevated as close to the inlet brine temperature as desired. The product water is then used to heat the countercurrent salt water in heat exchanger. The only heat requirement is to elevate the salt water from the heat exchanger outlet temperature to the evaporator inlet temperature. The elevation of temperature required is the sum of the temperature driving forces in the evaporator and the heat exchanger. In order to keep the heat requirement at a minimum, the temperature driving forces should be at their minimum. But the area required for heat transfer is inversely proportional to the temperature driving force for a heat exchanger. If the exchanger driving force is at its minimum, the area required would be infinite, and so would be the equipment cost. The temperature driving force in the evaporator, though not influencing the mass transfer at a given temperature, also affects the required evaporator area. This area increases to infinity as driving force decrease to the minimum, the boiling point elevation. Also at lower temperatures the area requirement increases rapidly. Therefore, any attempt to reduce the heat cost would increase the equipment costs of either evaporator or heat exchanger, or both, and decreasing equipment costs would

TABLE 4.4 COST COMPARISON

BETWEEN SIMILAR DESALINATION PROCESSES

	Distillation* thin-film 22-effect	Distillation** flash, 84-stage	Electro-dialysis***	EVAPORATION THROUGH MEMBRANES Method I	Method II
Capacity, gal.	50,000,000	50,000,000	10,000,000	10,000,000	10,000,000
Maximum temperature, ^{o}F	252	245	50	500	500
Power consumed Kw-hr./M gal.	2.61	7.03	9-12	5.59	5.59
Fuel, BTU/M gal.	0	0	0	1.091×10^{6}	0.798×10^{6}
Steam, BTU/M gal.	0.529×10^{6}	0.467×10^{6}	0	0	0
Cost/M gal.	$0.314	0.291	0.34-0.36	0.596	0.472

* Adapted from (20)

** Adapted from (21)

*** Adapted from (22)

increase heat cost. This is the main restriction on reducing the water cost.

A combined membrane evaporator-heat exchanger will partially overcome this disadvantage. In this system, Method II, the product water on shell side absorbs the heat of condensation and conduction from the concurrent, hot brine in the membrane tube. But the absorbed heat is not used to elevate the product water temperature. The product water transfers the absorbed heat plus its own sensible heat to the countercurrent feed salt water in the heat exchanger tubes. In this manner, the feed salt water can be heated to a temperature close to product water temperature at the heated salt water outlet end. The temperature distribution curves in Figure 4.8 through Figure 4.15 indicate that the temperature driving force across the membrane increases, but that across the heat exchanger tubes decreases only slightly as evaporating solution temperature decreases. This characteristic favors the operation at lower temperatures, by increasing the temperature difference for evaporation at lower temperatures. An important improvement of Method II is that the temperature driving force across the membrane can be increased to reduce the large requirement of membrane area and the resulting sharp increase of costs of membrane and evaporator at lower temperatures. The slight decrease of temperature driving force across the exchanger tubes does not affect the water cost very much because the heat transfer cost is smaller than mass transfer cost in the low temperature part of the main evaporator. The heated salt water temperature from the tubes controls not only the cost of heat but also

that of the evaporator. A high outlet temperature from the tubes would reduce heat costs, but it would increase the evaporator costs due to the small temperature driving forces available for mass and heat transfer. The cost study was, therefore, focused on optimizing this outlet temperature with maximum practical perimeters in an evaporator.

The additional advantages of Method II are that the equipment cost can be reduced by combining membrane and heat exchanger in one apparatus and that corrosion can be avoided by using the product water on the shell side, with saline water inside the membrane channel and inside exchanger tubes.

D. Cost Comparison Between Similar Processes:

A crude cost comparison is given in Table 4.4. For distillation processes, the water cost from different sources has been evaluated in the range of 25-35 cents per 1000 gallons. It should be emphasized here that the larger plant size of distillation processes requires lower capital and labor costs and sometimes lower fuel and power costs per unit of fresh water produced. In addition, the location and site of the plant have also been properly selected in order to minimize the capital and operating costs. Although Method II has been considered as having several advantages over similar distillation processes, the higher cost of Method II shows that the present calculated performance of Method II has not been brought to a competitive level. The method requires improvements on the membrane properties or process design before it can be recommended for desalination.

A major reason for the higher costs in this process are the lower heat transfer coefficients assumed for heat exchange, 200 BTU/hr. ft.^{2} ^oF, which is due to the fact that in other evaporation process boiling liquids and condensing vapors are heat exchange fluids while in this case the heat exchange fluids are liquid-liquid.

However, it should be emphasized that all of these methods are still in an experimental stage of development and improvements appear to be probable in the future. The least developed of the methods compared is the evaporation through membrane process; and on this basis the method appears to be most susceptible to improvement, either in membranes or operation methods.

For electrodialysis process, the amount of electric current required and resulting demineralization cost is in proportion to the amount of salt to be removed. Consequently, this process is favored for brackish waters containing considerably less dissolved salts than sea water. In this brackish-water region of about 200 ppm, the water cost is shown in Table 4.4 for a 10 MGD plant demineralizing 3000 ppm water to the 500 ppm level. Demineralizing sea water to a 500 ppm potable water requires a total cost of $0.48 per 1000 gallons based on cost optimization methods and idealized costs (22). With equal water cost, Method II is believed capable of producing water of a better quality than 500 ppm (18).

For the reverse osmosis process, the demineralization cost varies widely from $0.25 to 1.33 per 1000 gallons of product

water. The high membrane replacement cost leaves this process in an under-developed state, with considerable opportunity for improvement.

V. CONCLUSIONS

This study of the economics and design of two new desalination processes utilizing evaporation through porous, water-repellent membranes has led to the following conclusions:

1. The combined membrane evaporator-heat exchanger system designed in Method II appears more economically feasible for desalination than the separate system in Method I. The optimum cost is $0.472 per 1000 gallons of water product for Method II, and $0.596 for Method I.

2. Higher temperatures and pressures up to the maximum appears to be the most economically feasible range of operation.

3. The combined membrane evaporator-heat exchanger system in Method II increases the temperature driving force for mass transfer at lower temperatures and favors the operation down to the ambient temperature range.

4. As evaluated from the present knowledge of membrane performance, the desalination cost of sea water is apparently higher than that of other successful distillation processes, but is competitive with other membrane processes and appears likely to become more competitive in the future.

5. Similar to other membrane processes, membrane performance does control, along with heat exchanger performance, the evaluation of the present process.

VI. RECOMMENDATIONS

The following recommendations are suggested for future investigation in this field:

1. The relationship between mass transfer coefficients and flow, countercurrent or concurrent, should be exactly established for various kinds of membranes at different flow rates, salt concentrations, temperatures and pressures.

2. A more detailed cost estimate should involve mass transfer coefficients and heat transfer coefficients as a function of pressure, temperature, and flow rates when such relationships become available.

3. A pilot plant should be prepared to verify the optimum set of conditions.

4. Further experimental work on mass transfer coefficients for membranes of this type should be concentrated in the high temperature and pressure regions necessary for optimum results, and further design consideration should be given to reducing the cost of heat transfer and to reducing mass transfer costs at lower temperatures.

5. Further improvement of membranes for desalination purpose should be investigated.

VII. APPENDICES

APPENDIX A

ACTIVITY OF WATER IN SALT SOLUTION

The activity of the aqueous salt solution (7% by weight salt) was estimated from data in Chemical Engineering Handbook, 4th ed., p. 11-30, at several temperatures as follows:

$$(a_w)_{T+\Delta T} = \left(\frac{f_w}{f^o}\right)_{T+\Delta T} = \frac{(f^o)_T}{(f^o)_{T+\Delta T}}$$

where $(a_w)_{T+\Delta T}$ = activity of water in solution at a temperature of $T+\Delta T$,

(f_w) = fugacity of water in salt solution,

(f^o) = fugacity of pure water, assumed equal to vapor pressure,

T = temperature, ^oF

ΔT = boiling point elevation.

This activity of water remains approximately constant at 0.96 in the range investigated.

APPENDIX B

VAPOR PRESSURE OF WATER

The vapor pressure of water can be expressed as a function of temperature as follows (24):

$$\log \frac{P_c}{P} = \frac{X}{T} \left(\frac{a + bX + cX^3}{1 + dX} \right)$$

where P = the vapor pressure in atm.

P_c = 218.167 atm.

T = absolute temperature in ^{o}K.

T_c = 647.27

$X = T_c - T$

a = 3.2438

$b = 5.8683 \times 10^{-3}$

$c = 1.1702 \times 10^{-8}$

$d = 2.1878 \times 10^{-3}$

APPENDIX C

LATENT HEAT OF WATER

In the range of 50 to 600^{o}F, the latent heat of vaporization for water can be estimated from the following equation within 2% error:

$$h = 1074.8 - 0.290 \times T - 8.64 \times 10^{-4} \times T^{2}$$

where h is the latent heat of water in BTU per lb., and T is water temperature in ^{o}F.

This equation was fitted to the steam tables of Keenan and Keyes (24).

APPENDIX D

COMPUTER PROGRAM

The program used for the computations described in this thesis is given in this appendix. The program was written in WATFOR language and was run in IBM 360 system.

```
/WAT4  CN12001,TIME=5,PAGES=50                YANG K H                08/17/67  16495
C      METHOD I
C      CALCULATE ALL NECESSARY ITEMS FOR EVAPORATOR DESIGN
C      COST ESTIMATION
      DIMENSION P(450),H(450),PA(450),TF(450)
      DIMENSION FC(450),FA(450)
      B1=3.2438
      B2=5.8683E-3
      B3=1.1702E-8
      B4=2.1878E-3
      CP=1.
      U=13.
      CM=0.3
      TF(1)=75.
      AC=0.96
      DTM=5.
      DO 10 I=1,90
      TC=0
      TC=TC+(TF(I)-32.)/1.8
      X=374.11-TC
      FXX=2.303*X*(B1+B2*X+B3*X**3)/((TC+273.16)*(1.+B4*X))
      P(I)=218.18*29.92*EXP(-FXX)
      PA(I)=AC*P(I)
      H(I)=1074.8-0.29*TF(I)-8.64*(TF(I)/100.)**2.
   10 TF(I+1)=TF(I)+5.
      DO 60 K=1,3
      DT=5.
      DO 50 L=1,5
      C=1000.
      A=0.
      DC=0.
      IF(K-2)21,22,23
   21 N=76
      J=86
      COMO=3.9E+04
      COPM=1.5E+05
      COEG=1.2E+05
      CPO=387.
      GO TO 20
   22 N=46
      J=56
      COPM=2.6E+4
      COEG=2.4E+4
      COMO=1.0E+4
      CPO=71.6
      GO TO 20
   23 N=18
```

```
      J=28
      COPM=0.
      COMO=0.
      COEG=0.
      CPO=0.
   20 DA=0.
      JKL=J-L
      FC(J)=CP/(H(J)+U*DT/(CM*(PA(J)-P(JKL))))
      DC=DC+C*FC(J)*DTM
      FA(J)=C*CP/(CM*(PA(J)-P(JKL))*H(J)+U*DT)
      DA=DA+FA(J)*DTM
      C=1000.-DC
      A=A+DA
      DTEX=10.
      IF(J-6)13,12,13
   13 IF(J-N)11,12,11
   12 N=N-10
      DO 30 I1=1,4
      COFT=A/DC*6.95E+06
      COBR=(DT+DTEX)/DC*4.16E+06
      TEO=500.-DT-DTEX
      COE1=(TEO-TF(J))/DTEX/DC*3.8E+07
      IF(J-26)16,16,17
   17 COE2=(DT+DTEX)/DTEX/DC*3.8E+07
      COE3=(TF(J)-DT-DTEX-70.)/DTEX*6.94E+04
      GO TO 18
   16 COE2=0.
      COE3=(TF(J)-70.)/DTEX*6.94E+04
   18 COEX=COE1+COE2+COE3
      PIE=COPM+COMO+COEG+COEX+COFT+COBR
      TPI=1.813*PIE+2.818E+05
      CMB=A/DC*1675.
      CHT=(DT+DTEX)/DC*955*24
      TOP=CPO+CMB+CHT+PIE*5.90E-4+380.2
      CMG=TOP*1.E-4
      WCP=60.*TOP
      TTCP=TPI+WCP
      CCGD=TTCP*1.E-7
      WRITE(3,190)TF(J),DT,DTEX,C,DC,A
      WRITE(3,240)
      WRITE(3,200)COE1,COE2,COE3,COBR
      WRITE(3,210)
      WRITE(3,200)COFT,COEX,PIE,TPI
      WRITE(3,220)
      WRITE(3,200)CMB,CHT,TOP
      WRITE(3,230)
      WRITE(3,200)CMG,WCP,TTCP,CCGD
```

```
 91        30  DTEX=DTEX+10.
 92        11  J=J-1
 93            IF(J-6)50,6,6
 94         6  GO TO 20
 95        50  DT=DT+5.
 96        60  CONTINUE
 97       190  FORMAT(//6X,'TF(J)=',F5.0,5X'DT=',F4.0,5X,'DTEX=',F4.0,3F10.2)
 98       200  FORMAT(6E18.8)
 99       210  FORMAT(7X'COST FILTER',4X'COST EXCHANGER'9X'TOTAL PIE'12X'INVEST')
100       220  FORMAT(8X'COST MENBR',9X'COST HEAT',8X'OPER. COST')
101       230  FORMAT(9X'COST/MGAL'8X'WORK CAPTL'8X'TOTAL CAP.'10X'COST/GAL')
102       240  FORMAT(11X'COST F1',11X'COST E2',11X'COST E3',10X'COST BR.')
103            STOP
104            END
```

```
/DATA
```

```
/WAT4  CN120012,TIME=9,PAGES=50          YANG K H          08/17/67  16361
C      METHOD II
C      CALCULATE ALL NECESSARY ITEMS FOR EVAPORATOR DESIGN
C      COST ESTIMATION
       B1=3.2438
       B2=5.8683E-3
       B3=1.1702E-8
       B4=2.1878E-3
       CP=1.1
       V=13.
       CM=0.3
       AC=0.96
       TO=.5
       DT=5.
       PE=2.093
       PM=1.088
       DTM=2.
       DO 30 J=1,5
       DO 25 I=1,3
       IF (I-2)15,16,17
    15 TS=500.
       COPM=1.5E+05
       COMO=3.9E+04
       COEG=1.2E+05
       CPO=387.
       GO TO 18
    16 TS=350.
       COPM=2.6E+4
       COEG=2.4E+4
       COMO=1.0E+4
       CPO=71.6
       GO TO 18
    17 TS=212.
       COPM=0.
       COMO=0.
       COEG=0.
       CPO=0.
    18 TC=TS-DT
       TS2=TC-DT
       X=0.
       WS2=1000.
       WC=0.
       WS=1000.
       T1=TS
       TC1=(T1-32.)/1.8
       X1=374.11-TC1
       FXX1=2.303*X1*(B1+B2*X1+B3*X1**3)/((TC1+273.2)*(1+B4*X1))
```

```
      P1=218.18*29.92*EXP(-FXX1)
      H=1074.8-0.29*T1-8.64*(T1/100.)**2
      T2=TC
      TC2=(T2-32.)/1.8
      X2=374.11-TC2
      FXX2=2.303*X2*(B1+B2*X2+B3*X2**3)/((TC2+273.16)*(1+B4*X2))
      P2=218.18*29.92*EXP(-FXX2)
      CK = CM*PM*(P1-P2)/(T1-T2)
      E=1.0+.00267*TS
      U=V*PM
      UE = 200.*PE
      TC=(UE*TS2+H*CK*(TS-E)+U*TS)/(CK*H+U+UE)
      TCH=500.
    6 WRITE(3,100)WS,TS,TC,TS2,X,PE,DT
      IF(TS-150.)11,11,12
   11 TCH=TCH-10.
      GO TO 13
   12 TCH=TCH-50.
   13 IF(TCH-450.)7,8,7
    7 COEE1=(2.0+1.840)*X/(WC*24.)*8.33E+07
      COEE2=0.
      COEE3=0.
      CMB1=PM*X/WC*1675.
      CMB2=0.
      COBR=2.*DT/WC*4.16E+06
      COEE3=(11.*(TS2-75.)/((WS*TS+WC*TC)/1000.-TS2))/(WC*24.)*8.33E+7
      IF(COEE3)61,62,62
   61 COEE3=0.
   62 COEE=COEE1+COEE2+COEE3
      PIE=COPM+COMO+COEG+COEE+COBR
      TPI=1.813*PIE+2.818E+05
      CMB=CMB1+CMB2
      CHT=2.*DT/WC*955.*24.
      TOP=CPO+CMB+CHT+PIE*5.90E-4+380.2
      CMG=TOP*1.E-4
      WCP=60.*TOP
      TTCP=TPI+WCP
      CCGD=TTCP*1.E-7
      WRITE(3,250)
      WRITE(3,200)COEE1,COEE2,COEE3
      WRITE(3,240)
      WRITE(3,200)COBR,COEE,PIE,TPI
      WRITE(3,220)
      WRITE(3,200)CMB,CHT,TOP
      WRITE(3,230)
      WRITE(3,200)CMG,WCP,TTCP,CCGD
    8 T1=TS-.5*DTM
```

```
      TC1=(T1-32.)/1.8
      X1=374.11-TC1
      FXX1=2.303*X1*(B1+B2*X1+B3*X1**3)/((TC1+273.2)*(1+B4*X1))
      P1=218.18*29.92*EXP(-FXX1)
      H=1074.8-0.29*T1-8.64*(T1/100.)**2
      T2 =TC-.5*DTM
      TC2=(T2-32.)/1.8
      X2=374.11-TC2
      FXX2=2.303*X2*(B1+B2*X2+B3*X2**3)/((TC2+273.16)*(1+B4*X2))
      P2=218.18*29.92*EXP(-FXX2)
      CK = CM*PM*(P1-P2)/(T1-T2)
      E=1.+.00267*TS
      DC=CP*WS*DTM/(H+.5*CP*DTM+U*(TS-TC)/(CK*(TS-TC-E)))
      DX=DC/(CK*(TS-TC-E))
      DTC=(-CK*DX*(H+CP*(TS+.5*DTM-TC))*(TS-TC-E-.5*DTM)-U*DX*(TS-TC-
     1.5*DTM)+UE*DX*(TC-TS2+.5*DTM))/(.5*CK*DX*(H+CP*(TS+.5*DTM-TC))+.5
     1*U*DX+CP*WC+UE*DX*.5)
      DX = DC/(CK*(TS-TC-.5*DTM + .5*DTC-E))
      TS=TS-DTM
      TC = TC-DTC
      TS2 = TS2 -(WS*DTM + DC*(TS- TC) + WC*DTC)/WS2
      WS=WS-DC
      WC=WC+DC
      X=X+DX
      IF (TS-TCH) 9,9,8
    9 IF(TS-80.)25,25,6
   25 CONTINUE
   30 DT=DT+5.
      STOP
  100 FORMAT(//7F12.3)
  200 FORMAT(6E18.8)
  220 FORMAT(8X'COST MENBR',9X'COST HEAT',8X'OPER. COST')
  230 FORMAT(9X'COST/MGAL'8X'WORK CAPTL'8X'TOTAL CAP.'10X'COST/GAL')
  240 FORMAT(11X'COST BR'11X'COST FE'9X'TOTAL PIE'12X'INVEST')
  250 FORMAT(13X'COEE1',13X'COEE2',13X'COEE3')
      END

/DATA
```

APPENDIX E

COST RESULTS OF METHOD I

Following the standardized procedure (17) for first estimate of desalination cost, the cost results of Method I with sea water from $500^{o}F$ to $100^{o}F$ in the evaporator and with $D_T = 15^{o}F$ and $D_{TEX} = 20^{o}F$ are listed below.

Items of Cost Information.

The following is a summary of information items to be supplied as part of the cost estimate.

(a) The estimate is for 10 million gallons of product water per stream day.

(b) Twenty million gallons of sea water feed per stream day.

(c) Ten million gallons of waste concentrated sea water produced per stream day.

(d) The selection of materials of construction and plant equipment is based on a twenty year plant life.

(e) 97.2 kilowatt hours consumed per stream day.

(f) 9,300 millions of BTU consumed per stream day.

(g) Total operating force: eight.

(h) Itemized list of special equipment, standard engineering equipment, including standby units and the installed cost of each item.

(1) Capital Costs

(i) Essential plant costs:

1.	Evaporator-condenser	$2,606,000
2.	Heat-exchanger	2,122,000
3.	Heater	424,000
4.	Pump	150,000
5.	Engine	120,000
6.	Motor	39,000
	Total PIE (Principal Items of Equipment) installed	$5,461,000
7.	Erection and assembly of plant - 30% of PIE	1,638,000
8.	Instruments - 4% of PIE	218,000
	Total Essential Plant Costs (1 through 8)	$7,318,000

(ii) Other plant costs:

9.	Raw water supply $5 per 1000 gallons of feed water per stream day	$ 500,000
10.	Product water storage (10 days) at $10 per 1000 gallons product water per stream day	$ 500,000
11.	Service facilities and buildings 10% of PIE for plants of 10 million gal/day capacity	$ 546,000
12.	Contingencies - 10% of total of above 11 items	$ 806,000

13. Engineering - 10% of above 12 items . . . . . $ 887,000

14. Interest on investment during construction - 4% of plant investment (sum of above 13 items . . . . . $ 390,000

15. Site $3 per 1000 gallons of product water per stream day. 30,000

Total Plant Investment (Sum of Above 15 Items). . $1,018,000

(2) Working Capital

60 days product at the total operating cost . . . $ 416,000

Total capital costs (plant investment plus working capital). $10,593,000

Capital cost per gallon per day of production . .$ 1.059

(3) Operating Costs

(i) Essential operating costs (Basis, one stream day and 330 operating days per year)

		Cost Per Stream Day
1.	Fuel at 25 cents per 1,000,000 BTU.	$ 2,334
2.	Electric power - continuous demand below 100,000 Kw 7 mills per KWH.	$ 387
3.	Membrane at 5.8 cents per sq. ft. for 4 months life	$ 628
4.	Supplies and maintenance materials - 0.0015 percent of total plant investment	$ 153
5.	Operating labor, 5% of above 4 items for plants of 10,000,000 gal/stream day	$ 175

6.	Maintenance labor, 0.0015 percent of total plant investment.	$ 153
7.	Payroll extras, 15 percent of items 5 and 6 .	$ 50
	Total essential operating costs (sum of above 7 items	$ 3,879
(ii)	Other operating costs	
8.	General overhead and administrative overhead - 30 percent of items 5, 6 and 7 . .	$ 113
9.	Amortization - 0.0224 percent of total plant investment	$ 2,280
10.	Taxes and insurance, 0.006% of total plant investment.	$ 611
11.	Interest on working capital. Calculate as 0.00725 times the sum of the above 10 items	$ 50
	Total operating costs for one stream day (sum of above 11 items)	$ 6,932
	From the total operating costs per stream day calculate the cost per 1000 gallons of product water	$ 0.693

APPENDIX F

COST RESULTS OF METHOD II

Following the procedure for water cost estimates given by the Office of Saline Water (17), the cost results of Method II with $T_{s1} = 500^{o}F$, $T_{s2} = 130^{o}F$ and $T_{s31} = 470^{o}F$ are listed below.

Items of Cost Information

The following is a summary of information items to be supplied as part of the cost estimate:

(a) The estimate is for 10 million gallons of product water per stream day.

(b) Twenty million gallons of sea water feed per stream day.

(c) Ten million gallons of waste concentrated sea water produced per stream day.

(d) The selection of materials of construction and plant equipment is based on a twenty year plant life.

(e) 97.2 kilowatt hours consumed per stream day.

(f) 7,750 millions of BTU consumed per stream day.

(g) Total operating force: eight.

(h) Itemized list of special equipment, standard engineering equipment, including standby units and the installed cost of each item.

(l) Capital Costs

(i) Essential plant costs:

1. Evaporator-condenser $2,370,000

2.	Heat exchanger	$ 42,400
3.	Heater	352,000
4.	Pump	150,000
5.	Engine	120,000
6.	Motor	39,000
	Total PIE (principal items of equipment) installed	$3,073,400
7.	Erection and assembly of plant - 30% of PIE	$ 922,100
8.	Instruments - 4% of PIE	$ 123,000
	Total Essential Plant Costs (1 through 8)	$4,119,000
9.	Raw water supply $5.0 per 1000 gallons of feed water per stream day	$ 500,000
10.	Product storage (10 days) at $10 per 1000 gallons product water per stream day	$ 500,000
11.	Service facilities and buildings 10% of PIE for plants of 10 million gal/day capacity	$ 307,000
12.	Contingencies - 10% of total of above 11 items	$ 427,000
13.	Engineering - 10% of above 12 items	$ 509,000
14.	Interest on investment during construction - 4% of plant investment (sum of above 13 items	$ 224,000
15.	Site $3 per 1000 gallons of product water per stream day	$ 30,000

Total Plant Investment (Sum of Above 15 Items). . $5,852,000

(2) Working Capital

60 days product at the total operating cost . . . $ 286,000

Total capital costs (plant investment plus working capital). $6,138,000

Cost per gallon per day of production . . . . . $ 0.614

(3) Operating Costs

(i) Essential operating costs (Basis, one stream day and 330 operating days per year):

		Cost Per Stream Day
1.	Fuel at 25 cents per 1,000,000 BTU. . . . . $	1,942
2.	Electric power - continuous demand below 100,000 Kw 7 mills per KWH. . . . . $	387
3.	Membrane at 5.8 cents per sq. ft. for 4 months life. . . . . $	324
4.	Supplies and maintenance materials - 0.0015 percent of total plant investment . . . . . $	87.8
5.	Operating labor, 5% of above 4 items for plants of 10,000,000 gal/stream day . . . $	137
6.	Maintenance labor, 0.0015 percent of total plant investment. . . . . $	87.8
7.	Payroll extras, 15 percent of items 5 and 6 . $	33.7

Total essential operating costs (sum of above 7 items . . . . . $ 2,999

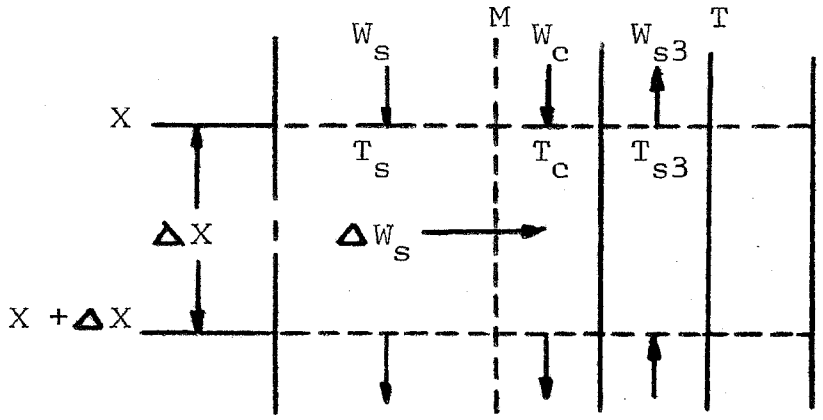
(ii) Other operating costs:

8.	General overhead and administrative overhead - 30% of items 5, 6 and 7. . . . . $	77.6
9.	Amortization - 0.0224 percent of total plant investment. . . . . $	1,311
10.	Taxes and insurance, 0.006% of total plant investment. . . . . $	351
11.	Interest on working capital. Calculate as 0.00725 times the sum of the above 10 items . $	34.1
	Total operating costs for one stream day (sum of the above 11 items . . . . . $	4,773
	From the total operating costs per stream day calculate the cost per 1000 gallons of product water.$	0.477

APPENDIX G

DERIVATION OF DIFFERENTIAL EQUATION FOR MASS AND HEAT BALANCE OF METHOD II

Differential equations describing the simultaneous mass and heat transfer of Method II are derived below.



Schematic Diagram for Mass and Heat Transfer of Method II.

Consider a small section of evaporator length, ΔX, as a system.

Mass balance for the system.

$$(W_s)_x - (W_s)_{x+\Delta x} + (W_c)_x - (W_c)_{x+\Delta x} - (W_{s3})_x + (W_{s3})_{x+\Delta x} = 0 \qquad \text{(i)}$$

For constant W_{s3}, equation (i) becomes

$$(W_s)_x - (W_s)_{x+\Delta x} + (W_c)_x - (W_c)_{x+\Delta x} = 0 \qquad \text{(ii)}$$

Mass transferred through the membrane

$$-\Delta W_s = K_m P_m (aP_A - P_{A2}) \Delta X \qquad \text{(iii)}$$

Heat balance for the system by taking zero reference temperature and assuming that specific heats of salt water and fresh water are equal and constant.

$$(C_p W_s T_s)_x - (C_p W_s T_s)_{x+\Delta x} + (C_p W_c T_c)_x - (C_p W_c T_c)_{x+\Delta x} - (C_p W_{s3} T_{s3})_x \; (C_p W_{s3} T_{s3})_{x+\Delta x} = 0 \quad \text{(iv)}$$

Heat balance for evaporating salt water stream, W_s.

$$(C_p W_s T_s)_x - (C_p W_s T_s)_{x+\Delta x} + (h + C_p \overline{T}_s) W_s - U P_m (\overline{T}_s - \overline{T}_c) \Delta X = 0 \quad \text{(v)}$$

Heat balance for heated salt water stream, W_{s3}.

$$- (C_p W_{s3} T_{s3})_x + (C_p W_{s3} T_{s3})_{x+\Delta x} + U_E P_E (\overline{T}_s - \overline{T}_c) = 0 \quad \text{(vi)}$$

Divide equation (ii) through equation (vi) by ΔX and then take limit as ΔX approaches zero. Then,

$$\frac{dW_s}{dX} + \frac{dW_c}{dX} = 0 \quad \text{(vii)}$$

$$- \frac{dW_s}{dX} = +K_m P_m (aP_A - P_{A2}) \quad \text{(viii)}$$

$$C_p \frac{d(W_s T_s)}{dX} + C_p \frac{d(W_c T_c)}{dX} - C_p W_{s3} \frac{dT_{s3}}{dX} = 0 \quad \text{(ix)}$$

$$C_p \frac{d(W_s T_s)}{dX} - (h + C_p T_s) \frac{dW_s}{dX} + U P_m (T_s - T_c) = 0 \quad \text{(x)}$$

$$C_p W_{s3} \frac{dT_{s3}}{dX} + U_E P_E (T_c - T_{s3}) = 0 \quad \text{(xi)}$$

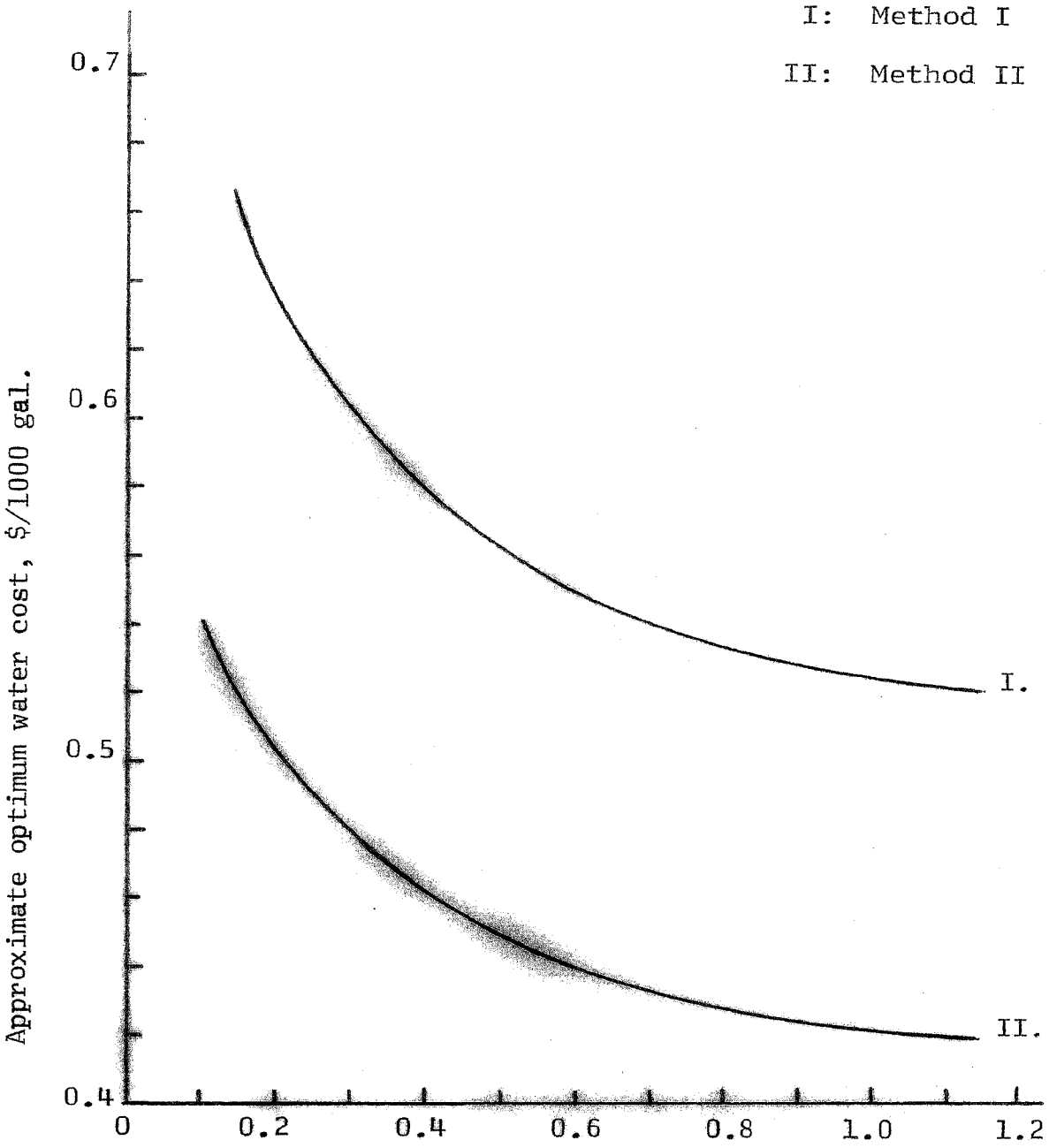
Another mass relationship is

$$W_{s3} = W_s + W_c \qquad \text{(xii)}$$

In the above equations, h, P_A and P_{A2} are complicated functions of temperature, a is a function of temperature and salt concentration, and the unknown variables to be solved for are W_s, W_c, W_{s3}, T_s, T_c, T_{s3} and X.

Theoretically, the above differential system (six equations) can be solved by expressing the six unknown variables each in terms of X. But the solutions are so complicated that it can not be obtained exactly by direct integration. Finally, the system is solved by means of difference equations. Based on equations (ii), (iii), (iv), (v), (vi) and (xii), and reasonable assumptions, the difference equations have been derived and discussed in the Design Equation of Method II.

APPENDIX H



Approximate Optimum Water Cost vs. Overall Mass Transfer Coefficient, K_m, with U = 13.0 BTU/hr.ft.^{2o}F

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